



INVESTMENT AND OPERATING COST STUDY
SULPHUR RECOVERY FACILITIES
NATURAL GAS PROCESSING INDUSTRY
(ENVIRONMENT CANADA)

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1.0 INTRODUCTION

This study provides capital investment and operating costs of Sulphur Recovery Facilities inclusive, Tail Gas Recovery systems and Stack Gas Emission Treatment in the natural gas processing industry.

The capital investment and operating costs are assessed for total sulphur recovery facilities. Feed stock to these facilities is acid gas. The sulphur recovery facilities requirements are based on three ratios of H_2S to CO_2 ratio in acid gas, i.e., 15/85, 50/50 and 80/20. This study covers a capacity range of 10 tonnes/day to 5000 tonnes/day of sulphur in the acid gas.

The capacity range of 10 tonnes/day to 5000 tonnes/day is further broken down into two groups. The first group contains Sulphur Plants from 10 tonnes/day to 100 tonnes/day. In this group, the Sulphur Recovery Unit is a Claus process plant and incinerator. Sulphur recovery of 97% can be achieved in 80/20 H_2S/CO_2 in 3 stage Claus process plant. A Tail Gas Plant, CBA or Sulfreen is always required for cases 15/85 or 50/50 H_2S/CO_2 cases to meet the 97% sulphur recovery efficiency. A separate installed cost for 3 stage Claus plant for 80/20 case and a combined installed cost for the two stage Claus plant and the Tail Gas Clean-up unit is provided. It is assumed that all of the H_2S and sulphur compound in exit stream from the Tail Gas Clean-up Unit are oxidized to SO_2 in the incinerator at $538^\circ C$. Waste heat recovery is not considered for any case after incineration of the tail gas. A combined installed cost is given for the incinerator and the stack for all the cases.

The second group contains Sulphur Recovery Units of capacities from 100 to 5000 tonnes/day. This group also utilizes Claus process for sulphur recovery, supplemented with tail gas Clean-up Unit incineration of the tail gas and stack. The sulphur recovery efficiency of 98% or better, can be achieved in 2 or 3 stage Claus plant added

1.0 INTRODUCTION (cont'd)

with any one of the CBA, Sulfreen, SCOT, BSR/MDEA, BSR/Selectox or BSR/STRETFORD Tail Gas Clean Up Unit. In this group, also a combined installed cost for the sulphur Recovery Unit and a Tail Gas Clean-up Unit for each case is provided. A separate cost for incinerator and stack is given. As in the first group waste heat recovery from incinerated tail gas is not considered.

Plant capacities of 50 and 150 t/d for 15/85 H_2S/CO_2 ratio, 15 and 350 t/d for 50/50 H_2S/CO_2 ratio and 100, 400 and 1000 t/d for 80/20 H_2S/CO_2 ratio, are chosen for modeling utility balances and operating cost data. A breakdown of installed costs for the basic Claus plant, different Tail Gas plant, incinerator and stack is provided for these capacities.

2.0 STUDY BASIS

2.1 Feed Gas Composition

The feed gas composition for all the cases considered, are given in Table 2-1. The feed gas is assumed to enter the Sulphur Recovery Unit battery limit at 43°C and 170 kPa (ab.), saturated with water and contains 0.5 vol% methane (wet basis) for all the cases studied. The acid gas is assumed to be free of other contaminants like NH₃, COS, CS₂ and RSH. The acid gas battery limit conditions of 43°C and 170 kPa (ab.), are commonly available for the Sulphur Recovery Unit in gas plants and refineries.

TABLE 2-1

FEED GAS COMPOSITION

<u>H₂S/CO₂</u> <u>Ratio</u>	<u>15/85</u>	<u>50/50</u>	<u>80/20</u>
Case No.	1-8	9-16	17-29
Mole %	-	-	-
H ₂ S	14.15	47.16	75.45
CO ₂	80.17	47.16	18.87
CH ₄	0.50	0.50	0.50
H ₂ O	5.18	5.18	5.18
MW	41.12	37.8	35.03

2.0 STUDY BASIS (cont'd)

2.2 Plant Capacities

Investment cost data is provided for 10 to 100 tonnes per day and 100 to 5000 tonnes/day sulphur plants with various combinations of the tail gas clean-up units.

For operating cost purposes, the plant capacities of 50 and 150 tonnes per day for 15/85 H_2S/CO_2 ratio, 15 and 350 tonnes per day for 50/50 H_2S/CO_2 ratio and 100, 400 and 1000 tonnes per day for 80/20 H_2S/CO_2 ratio, are considered, as illustrated in Table 2-2. These plant capacities are chosen because they cover most of the sulphur plants with 15/85, 50/50 and 80/20 H_2S/CO_2 ratio in Canadian gas processing industries.

TABLE 2-2

PLANT CAPACITIES

SELECTED FOR OPERATING COST ESTIMATES

<u>Case No.</u>	<u>1-2</u>	<u>3-8</u>	<u>9-10</u>	<u>11-16</u>	<u>17</u>	<u>18-23</u>	<u>24-29</u>
H_2S/CO_2 Ratio	15/85	15/85	50/50	50/50	80/20	80/20	80/20
Sulphur t/d in feed gas	50	150	15	350	100	400	1000

2.0 STUDY BASIS (cont'd)

2.3 Sulphur Recovery Level

Sulphur recovery level is divided into two groups with respect to the plant capacity as listed in Table 2-3. A sulphur recovery level of 97% minimum for 10 to 100 tonnes per day and 98% minimum for 100 - 5000 tonnes per day are required*. A Claus plant shall be supplemented with a Tail Gas Clean-up Unit if the Claus plant alone is not sufficient to meet the required minimum recovery level. Table 2-3 also indicates various Tail Gas Clean-up units to be studied in various combinations with the Claus plant.

Table 2-3

<u>SULPHUR RECOVERY LEVEL</u>		
<u>Plant Capacity t/d</u>	<u>Minimum Sulphur Recovery Level %</u>	<u>Tail Gas Clean-up Unit Type</u>
10 - 100	97	CBA, Sulfreen,
100 - 5000	98	CBA, Sulfreen, SCOT, BSR /MDEA, BSR/Selec- tox or BSR/Stretford

* It is realized in some particular situations, the required recovery level may be higher.

2.0 STUDY BASIS (cont'd)

2.4 Other Requirements, Criteria and Assumption

(Following requirements, criteria and assumptions are based on Environment Canada letter 47SS.KE 145-2-0713 dated December 16th, 1982 from Mr. J.R. Chase to Mr. J. Newman of MHG International Ltd.).

- 1) Barometric pressure is 101.3 kPa (ab.).
- 2) Acid gas is available at 43°C and 69 kPa(g), saturated with water and contains 0.5 vol% methane (wet basis).
- 3) 4140 kPa(g), 345 kPa(g) and 105 kPa(g) steam is produced in waste heat boiler and sulphur condensers in the Claus Sulphur Recovery Unit. 345 kPa(g), 105 kPa(g) steam generation or preheating of BFW is carried out in the Tail Gas Clean-up Unit.
- 4) Split Claus process flow scheme is used for 15/85 H₂S/CO₂ ratio feed gas cases. Straight Claus through process scheme is applied to 50/50 and 80/20 H₂/CO₂ ratio cases.
- 5) Reheating of first condenser effluent is done using inline reheater, burning acid gas. 4140 kPa(g) steam is used for reheating feed acid gas to second or third stage of conversion.
- 6) Treated deaerated boiler feed water is available at 2205 kPa (ab.) and 110°C.
- 7) Cooling water is available at 29.5°C and maximum return temperature of 43.5°C is considered.

2.0 STUDY BASIS (cont'd)

2.4 Other Requirements, Criteria and Assumption (cont'd)

- 8) This study provides installed costs for 75m and 152m high stack for all the cases.
- 9) Exit stack gas assumed to contain 2.5% excess oxygen and temperature of the exit stack gas is assumed to be 538°C.
- 10) The components of investment and operating costs are based upon August, 1982 for plants located in Calgary, Alberta. The costs are upgradable to future dates through the use of such indices as labor costs, material cost and construction cost.
- 11) Investment on installed cost includes material, labor and off-side costs, fees, taxation and initial charge for chemical and catalysts.
- 12) Operating cost includes supervisory and operating personnel to cover continuous plant operation. Also included are maintenance, taxes, insurance, catalyst, chemicals, cooling water, energy consumption as well as overhead costs*. Unit costs required for determining operating cost and are defined in pertinent section on operating cost.
- 13) Credits are given for battery limit export items, steam, and elemental sulphur production.

Note: *Based on the final review meeting between Mr. J. Labuda and MHG on May 29th, 1983, overhead and indirect operating costs are excluded.

3.0 SUMMARY OF RESULTS

SO₂ emission data for a total of 29 selected cases are given in Table 3-1. This Table also indicates various combinations of tail gas clean-up units. Split flow Claus process flow scheme is most suitable for the feed gas with a H₂S/CO₂ ratio of 15/85 to achieve around 98% sulphur recovery in Claus plant supplemented with a tail gas clean-up unit.

Investment and operating data for all of the selected cases are provided in Section 7 and Section 8, respectively. This study represents that for 15/85 and 50/50 H₂S/CO₂ ratios in the feed gas, a tail gas clean-up unit should always be followed by a Claus plant to achieve a 97% or 98% sulphur recovery. In case of 80/20 H₂S/CO₂ ratio in the feed gas, a 97% sulphur recovery can be achieved with 3 a stage Claus plant. Therefore, for plants between 10 and 100 t/d where the required sulphur recovery is 97%, a tail gas clean-up unit is not required. For plant capacities more than 100 t/d, a tail gas clean-up unit is required.

Plant capacities of 50 and 150 t/d for H₂S/CO₂ ratio of 15/85, 15 and 350 for H₂S/CO₂ ratio of 50/50 and 100, 400 and 100 t/d for H₂S/CO₂ ratio 80/20 are chosen for modeling operating cost data. These plant capacities cover most of the operating plants.

Two stack heights of 76m and 152m are studied. Each of these is equipped with an incinerator to completely convert all sulphur containing compounds in the tail gas to SO₂. Capital cost data for combined incinerator and stack, for each case, are given in Figure 7.10 and 7.11.

3.0 SUMMARY OF RESULTS (cont'd)

Table 3-2 summarizes sulphur recoveries that can be achieved in various Claus plant. It also shows the overall sulphur recovery efficiency that can be reached with various combinations of the Claus plant and tail Gas Clean-Up Unit.

Daily quantities of H_2S and SO_2 at the entry and exit of the Tail Gas Clean-Up Unit and the incinerator outlet are also listed in Table 3-2.

TABLE 3-1

SO₂ EMISSION IN t/d AND ppav

Case No.	Sulphur Input t/d	Feed Gas H ₂ S/CO ₂ ratio	Sulphur Recovery Unit		Tail Gas Clean-Up Unit Type	SO ₂ Emission		Stack Height*	
			Stages	Type		t/d	ppav	m	
1	50	15/85	2	Sp. Flow Claus	CBA	2	2200	76	152
2	50	15/85	2	Sp. Flow Claus	Sulfreen	2	2200	76	152
3	150	15/85	2	Sp. Flow Claus	CBA	6	2200	76	152
4	150	15/85	2	Sp. Flow Claus	Sulfreen	6	2200	76	152
5	150	15/85	3	Sp. Flow Claus	SCOT	0.03	80	76	152
6	150	15/85	3	Sp. Flow Claus	BSR/MDEA	0.30	80	76	152
7	150	15/85	2	Sp. Flow Claus	BSR/Selectox	6.60	1730	76	152
8	150	15/85	3	Sp. Flow Claus	BSR/Stretford	-	-	30	30
9	15	50/50	2	Strght.Thru Claus	CBA	0.45	2790	76	152
10	15	50/50	2	Strght.Thru Claus	Sulfreen	0.45	2790	76	152
11	350	50/50	2	Strght.Thru Claus	CBA	10.5	2790	76	152
12	350	50/50	2	Strght.Thru Claus	Sulfreen	10.5	2790	76	152
13	350	50/50	3	Strght.Thru Claus	SCOT	.53	195	76	152
14	350	50/50	3	Strght.Thru Claus	BSR/MDEA	.53	195	76	152
15	350	50/50	2	Strght.Thru Claus	BSR/Selectox	10.5	2790	76	152
16	350	50/50	3	Strght.Thru Claus	BSR/Stretford	-	-	30	30
17	100	80/20	3	Strght.Thru Claus	Not Required	6	7560	76	152
18	400	80/20	2	Strght.Thru Claus	CBA	10.4	3280	76	152
19	400	80/20	2	Strght.Thru Claus	Sulfreen	10.4	3280	76	152
20	400	80/20	3	Strght.Thru Claus	SCOT	0.6	250	76	152
21	400	80/20	3	Strght.Thru Claus	BSR/MDEA	0.6	250	76	152
22	400	80/20	2	Strght.Thru Claus	BSR/Selectox	11.2	3535	76	152
23	400	80/20	3	Strght.Thru Claus	BSR/Stretford	-	-	30	30
24	1000	80/20	2	Strght.Thru Claus	CBA	26	3280	76	152
25	1000	80/20	2	Strght.Thru Claus	Sulfreen	26	3280	76	152
26	1000	80/20	3	Strght.Thru Claus	SCOT	1.5	250	76	152
27	1000	80/20	3	Strght.Thru Claus	BSR/MDEA	1.5	250	76	152
28	1000	80/20	2	Strght.Thru Claus	BSR/Selectox	28	3535	76	152
29	1000	80/20	3	Strght.Thru Claus	BSR/Stretford	-	-	30	30

* Two stack sizes of 76m and 152m height are studied in this report. The required height of stack for Sulphur Recovery Unit depends on location, surrounding and allowed ground level SO₂ concentration. The stack height of 152m is considered for installed cost of Sulphur Recovery Facilities in Section 7.

Table 3-2

Sulphur recovery efficiency, H ₂ S and SO ₂ quantities at entry and exit of tail gas clean-up unit and the incinerator outlet.												
Case No.	Sulphur Input t/d	H ₂ S/CO ₂ Ratio	Claus Sulphur Recovery Unit		Tail Gas Clean-up Unit Type	Overall Sulphur Recovery %	Tail Gas Clean-up Unit				Incinerator Outlet	
			No. Stages	%Sulphur Recovery			Entry***		Exit***		H ₂ S kg/d	SO ₂ kg/d
							H ₂ S kg/d	SO ₂ kg/d	H ₂ S kg/d	SO ₂ kg/d		
1	50	15/85	2	92	CBA	98.0	1450	1370	362	342	NII	2000
2	50	15/85	2	92	Sulfreen	98.0	1450	1320	362	342	NII	2000
3	150	15/82	2	92	CBA	98.0	4350	4110	1088	1028	NII	6000
4	150	15/82	2	92	Sulfreen	98.0	4350	4110	1088	1028	NII	6000
5	150	15/85	3	94	SCOT	99.9	3260	3080	61	0	NII	300
6	150	15/85	3	94	BSR/MDEA	99.9	3260	3080	61	0	NII	300
7	150	15/85	2	92	BSR/Selectox	97.8	4350	4110	1797	0	NII	6600
8	150	15/85	3	94	BSR/Stretford*	99.98	3260	3080	12	0	-	-
9	15	50/50	2	94	CBA	98.5	220	210	55	52.5	NII	450
10	15	50/50	2	94	Sulfreen	98.5	220	210	55	52.5	NII	450
11	350	50/50	2	94	CBA	98.5	5130	4900	1282.5	1225	NII	10500
12	350	50/50	2	94	Sulfreen	98.5	5130	4900	1282.5	1225	NII	10500
13	350	50/50	3	95	SCOT	99.9	4275	4080	80	0	NII	530
14	350	50/50	3	95	BSR/MDEA	99.9	4275	4080	80	0	NII	530
15	350	50/50	2	94	BSR/Selectox	98.5	5130	4900	2125	0	NII	10500
16	350	50/50	3	95	BSR/Stretford*	99.98	4275	4080	16	0	-	-
17**	100	80/20	3	97	Not Required	97.0	1740	-	-	-	NII	6000
18	400	80/20	2	94.5	CBA	98.7	8700	8590	2175	2150	NII	10400
19	400	80/20	2	94.5	Sulfreen	98.7	8700	8590	2175	2150	NII	10400
20	400	80/20	3	97	SCOT	99.9	4745	4680	90	0	NII	600
21	400	80/20	3	97	BSR/MDEA	99.9	4745	4680	90	0	NII	600
22	400	80/20	2	94.5	BSR/Selectox	98.6	8700	8590	3650	0	NII	11200
23	400	80/20	3	97	BSR/Stretford*	99.98	4745	4680	18	0	-	-
24	1000	80/20	2	94.5	CBA	98.7	21750	21470	5440	5370	NII	26000
25	1000	80/20	2	94.5	Sulfreen	98.7	21750	21470	5440	5370	NII	26000
26	1000	80/20	3	97	SCOT	99.9	11860	11710	226	0	NII	1500
27	1000	80/20	3	97	BSR/MDEA	99.9	11860	11710	226	0	NII	1500
28	1000	80/20	2	94.5	BSR/Selectox	98.6	21750	21470	5315	0	NII	28000
29	1000	80/20	3	97	BSR/Stretford*	99.98	11860	11710	45	0	NII	-

* BSR/Stretford cases do not required incinerator.

** In case 17 tail gas clean-up unit not required so entry to tail gas clean-up unit should be recognized as entry to incinerator.

*** Tail Gas Clean-up Unit entry and exit acid gas streams also contain COS, CS₂ and unrecovered element sulphur vapors (not shown above). All of these are assumed to be oxidized to SO₂ in incinerator.

4.0 SULPHUR RECOVERY PROCESSES

4.1 General

As described in Section 2, feeds to the Sulphur Recovery Units may contain 15%, 50% or 80% hydrogen sulphide. Selection of process and flow sheet is based primarily on acid gas composition. Straight through claus process is selected for feeds containing 50% or more hydrogen sulphide and split flow claus process is chosen for feeds containing 15% hydrogen sulphide.

4.2 Straight Through Claus Process

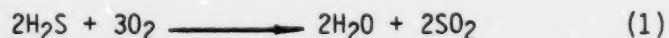
(Refer to Figure 4.1 for 2 stage and to Figure 4.2 for 3 stage Straight Through Claus Process).

In straight through claus process the Sulphur Recovery Unit produces sulphur according to the overall reaction.



This reaction takes place in the following steps:

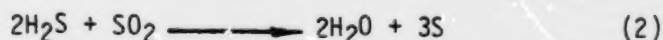
- a) Acid gas is burned in the reaction furnace with a carefully regulated amount of air so that one third of the H_2S in the acid gas is converted to SO_2 according to the chemical reaction:



4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.2 Straight Through Claus Process

- b) In the reaction furnace and waste heat boiler much of the H_2S and SO_2 in the gas stream reacts according to the following reaction:



As the gas cools, the elemental sulphur condenses out in the sulphur condenser and is removed.

- c) In the next stage, the effluent gas containing the reactants SO_2 and H_2S is reheated and passed through a catalyst bed which promotes reaction (2) shown above. The gas is again cooled to condense the sulphur product. The reheating, catalytic reaction and cooling steps are repeated to two or three stages depending upon the recovery level to be achieved. The sulphur plant gas effluent is either incinerated and discharged to atmosphere via a stack if the desirable recovery level is already met in sulphur recovery unit or sent to a tail gas clean-up unit to enhance the sulphur recovery to meet the emission requirement.

Process Air

Oxygen required for the combustion of acid gas in the reaction furnace (F-101) is provided by the air blower (K-101). The air flow rate to the burner of F-101 is set by flow ratio control, based on the acid gas flow rate to the Sulphur Recovery Unit. Control of the air flow rate to F-101 is further trimmed to maintain a 2 to 1 ratio of H_2S to SO_2 in the tail gas.

The air blower is driven by an electric motor.

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.2 Straight Through Claus Process (cont'd)

Reaction Furnace, Waste Heat Boiler and No. 1 Condenser

The acid gas enters the Sulphur Recovery Unit at 43°C. Free liquid (mostly sour water) is collected in the acid gas K.O. drum (V-101). The sour liquid is pumped intermittently out of the Sulphur Recovery Unit battery limit by pump (P-101).

The acid gas from V-101 is split into two streams. The first stream containing most of the net flow to SRU is mixed with combustion air and burned in the acid gas burner of F-101. The product gas is cooled in the Waste Heat Boiler (E-101). High pressure steam at 4140 kPa (g) is generated from cooling the process gas. No liquid sulphur is expected to be formed in the Waste Heat Boiler under normal operation.

No. 1 Reheater, No. 1 Converter, No. 2 Condenser

The second acid gas stream from V-101 contains part of the net acid gas flow to SRU and is also mixed with combustion air and burned in the acid gas burner of No. 1 Reheater (F-102). The hot acid gas effluent from F-102 is blended with the cold acid gas effluent from the E-102 to arrive at the appropriate temperature required for claus catalytic reaction in No. 1 Converter (V-102). In the converter (V-102) SO_2 and H_2S are converted to sulphur in the presence of alumina catalyst with an accompanying increase in temperature which will promote maximum conversion of COS and CS_2 .

No. 1 Converter effluent enters the No. 2 Condenser (E-103) where temperature is reduced by generation of 345 kPa (g) steam. Condensed sulphur from E-103 flows by gravity to the sulphur pit (T-101) via No. 2 Seal Pot (V-105).

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.2 Straight Through Claus Process (cont'd)

No. 2 Reheater, No. 2 Converter, No. 3 Condenser

Effluent acid gas from E-103 enters No. 2 Reheater (E-104). The gas is reheated to desired second stage converter inlet temperature, using 4140 kPa (g) saturated steam. The reheated acid gas enters the No. 2 Converter V-103 where SO_2 and H_2S again reacts to form sulphur.

The V-103 effluent flows to No. 3 Condenser (E-105) where the effluent temperature is brought down by generating 345 kPa (g) steam. Condensed sulphur flows by gravity to the sulphur pit (T-101) via Seal Pot (V-106).

No. 3 Reheater, No. 3 Converter, No. 4 Condenser

Similar steps of reheating acid gas, converting SO_2 and H_2S to sulphur and cooling converter effluent as in in proceeding paragraph are repeated in No. 3 Reheater (E-106), No. 3 Converter (V-107) and No. 4 Condenser (E-107). 4140 kPa (g) steam is used to reheat the E-105 acid gas effluent, and cooling of converter effluent is carried out by preheating the BFW in E-107. Condensed sulphur in E-107 flows to the Sulphur Pit (T-101) via Seal Pot (V-108).

Determination of the Number of Catalytic Converting Stages

Selecting the number of catalytic converting stages primarily depends on the desirable percentage of sulphur recovery and the capacity of sulphur plant in tonnes per day. The following data describes the number of stages required and the need of tail gas clean-up unit to achieve the required sulphur recovery in 50% and 80% H_2S cases in feed gas, using the straight through process.

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.2 Straight Through Claus Process (cont'd)

Determination of the Number of Catalytic Converting Stages (cont'd)

Plant Capacity T/D	In Feed Gas	Type	Basic Claus Plant		Tail Gas Clean-up Unit
			Anticipated Recov. %	Req'd Recov. %	
10-100	80/20	3STG.	97	97	Not Req'd
100-1000	80/20	3STG.	97	98	Req'd
10-100	50/50	3STG.	96.8	97	Req'd
10-100	50/50	2STG.	94.0	97	Req'd
100-350	50/50	2STG.	94.0	98	Req'd

In 10-100 tonnes/day, 50/50:H₂S/CO₂ cases a three stage Claus plant using straight through process gives 96.8% sulphur recovery which is slightly less than reequred 97%. Therefore a tail gas clean-up unit using a subdew point process is required.

For plants above 100 t/d, other tail gas clean-up processes are included.

In cases where a tail gas clean-up unit is not required tail gases from the sulphur recovery unit is incinerated and emitted to atmosphere through a stack.

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.2 Straight Through Claus Process (cont'd)

Incinerator and Stack

An incinerator (F-103) and a stack (U-101) are provided to convert the amount of unrecovered sulphur compounds to SO_2 and to dispense it into the atmosphere. This is achieved by burning fuel gas with air received from natural draft and then mixing the combustion products with the tail gas from E-107. The oxygen from natural draft, also oxidizes all the sulphur compounds to SO_2 under the incinerating conditions. The incinerated effluent gas leaves the stack at a minimum temperature of 530°C .

Boiler Feed Water

Boiler feed water enters the sulphur recovery unit at about 110°C . In three stage claus straight through process, BFW is preheated to 148°C in the No. 4 Condenser. The preheated BFW flows to No. 1, 2 and 3 Condensers for 345 kPa (g) steam production. Booster pump is provided for BFW to Waste Heat Boiler where high pressure 4140 kPa (g) steam is produced.

4.3 Split Flow Process

(Refer to Figure 4.3).

The split flow scheme is selected for feeds containing 15% hydrogen sulphide. The same reaction as in Equation (1) for the straight through process takes place in the split flow scheme in the reaction furnace. In this case only one third

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.3 Split Flow Process (cont'd)

of the total Sulphur Recovery Unit feed is fed to the reaction furnace with sufficient air to convert hydrogen sulphide to sulphur dioxide and hydrocarbon to carbon dioxide and water completely. The remaining two thirds of the acid gas feed, is blended with the waste heat boiler effluent prior to being charged to the 1st converter. Catalytic reaction which takes place in the converter is the same as the straight through process shown in Equation (2).

Process Air

Oxygen required for the combustion of acid gas to the furnace (F-101) is provided by the air blower (K-101). The air flow rate to the burner of F-101 is set by flow ratio control based on the acid gas flow rate to the furnace. Since only one third of the acid gas is fed to F-101 and the objective is to completely oxidize H_2S to SO_2 .

The air blower is driven by a electric motor.

Acid Gas, Reaction Furnace and Waste Heat Boiler

As in the straight through process, acid gas enters the SRU at $43^{\circ}C$. Free liquid (mostly sour water) is collected in acid gas K.O. drum (V-101). The sour water is pumped intermittently out of the SRU battery limit by pump (P-101). One third of the acid gas is fed to the reaction furnace (F-101) where acid gas is oxidized to SO_2 completely. The reaction furnace effluent is cooled in a two pass waste heat boiler. In the split flow scheme the second pass of waste heat boiler E-101 is a cooling pass. The second pass effluent is blended with part of the first pass acid gas effluent and remaining acid gas not charged to the F-101 to give the desired No. 1 converter inlet temperature.

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.3 Split Flow Process (cont'd)

No. 1 Converter, Gas/Gas Exchanger, No. 1 Condenser, No. 2 Converter and No. 2 Condenser

The waste heat boiler first and second pass effluent, along with the remaining acid gas to SRU, enters No. 1 Converter (V-102). In the converter (V-102), SO_2 and H_2S are converted to sulphur in the presence of alumina catalyst with an accompanying increase in temperature which will promote maximum conversion of COS and CS_2 .

The acid gas effluent from V-102 flows to No. 1 Condenser (E-103) via Gas/Gas Exchanger (E-102). The Gas/Gas Exchanger is used for reheating acid gas effluent from the No. 1 Condenser to the temperature required for No. 2 Converter (V-103). The acid gas effluent from V-103 is cooled in No. 2 Condenser (E-104).

Liquid sulphur streams from E-103 and E-104, are collected in the Storage Pit, T-101 via Seal Pots, V-104 and V-105 respectively. The acid gas effluent from E-104 flows to the Tail Gas Clean-up Unit to enhance the sulphur recovery level to 97% for plants of 10-10 t/d and above 98% for plants above 100 t/d.

Boiler Feed Water

Boiler feed water enters the Sulphur Recovery Unit at about 110°C . 4140, 345 and 105 kPa (g) steam is produced in the Waste Heat Boiler, No. 1 Condenser and in the No. 2 Condenser respectively. Booster pump is provided for BPW to Waste Heat Boiler where high pressure 4140 kPa (g) steam is produced.

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

4.4 Richards Sulphur Recovery Process (RSRP)

(Refer to Process flow Diagram 4.4).

Richards process carries out the conventional Claus reactions at elevated pressure in two steps. First catalytic furnace (the RSRP furnace, F-101) to oxidize H_2S , second a catalytic trickle bed reactor (the RSRP reactor, V-101) to react H_2S with SO_2 . The process is still at its early development stage. The process flow diagram and description are included for reference. Equipment design and cost estimates are not available.

Acid gas and air are compressed in compressors (K-101) and (K-102) before entering the RSRP furnace (F-101). The furnace contains a catalyst that promotes oxidation of H_2S to SO_2 . Cool liquid sulphur is pumped by pump (P-101) and sprayed on the catalyst in F-101 to maintain a required gas temperature during oxidation. Hot liquid sulphur, after passing through the catalytic bed of F-101, enters No. 1 Sulphur Cooler (E-102) and then No. 2 Sulphur Cooler (E-103) for heat recovery. High and medium pressure steam is produced in E-102 and E-103 respectively.

Acid gas effluent from F-101 enters catalytic reactor (V-101) which contains Claus catalyst to promote Claus reaction between H_2S and SO_2 in accordance with Equation (2) of Section 4.2. Here, cool liquid sulphur is also pumped by pump (P-101) and sprayed on the catalyst in V-101. The gases in V-101 are cooled by liquid sulphur. Hot sulphur, after passing through the catalytic bed, blends with the sulphur effluent from E-103 and then enters No. 3 Sulphur Cooler (E-104) for heat recovery. Low pressure steam is produced in E-104. The liquid sulphur from E-104 enters E-105 for further heat recovery to preheat boiler feed water.

4.0 SULPHUR RECOVERY PROCESSES (cont'd)

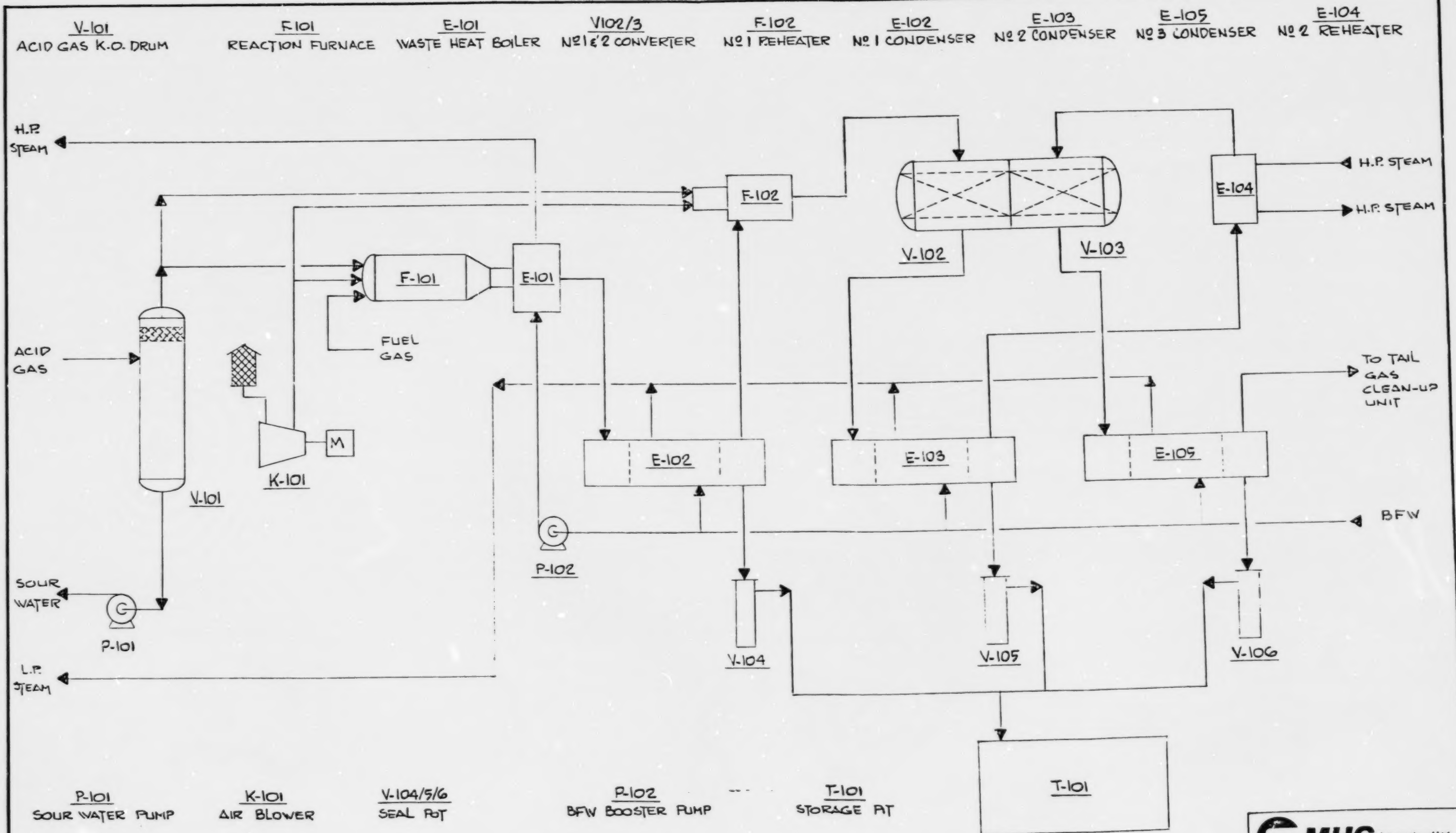
4.4 Richards Sulphur Recovery Process (RSRP) (cont'd)

Acid gas effluent from V-101 contains unreacted H_2S and sulphur vapors, which are recovered in the acid gas condenser E-101. Medium pressure steam is produced in E-101. Condensed sulphur from E-101 is collected in the Sulphur Storage Pit (T-101) along with sulphur from E-105.

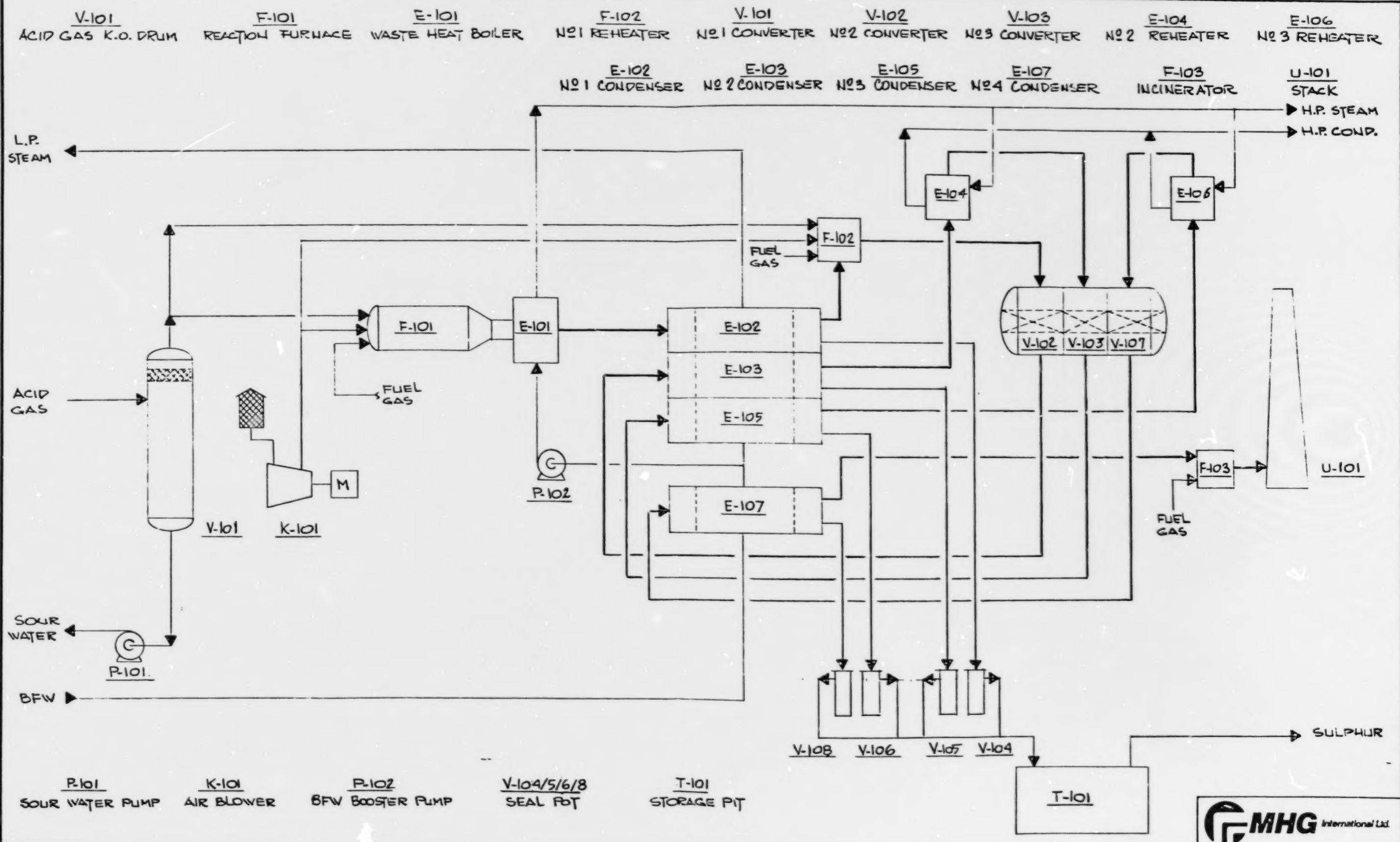
The sulphur recovery level is claimed to be as high as 98 to 99% depending on the number of beds in the RSRP reactor.

The tail gas from the unit is incinerated in the incinerator (F-102) with compressed air from K-102 and fuel gas so that all unreacted H_2S is converted to SO_2 for emission to atmosphere.

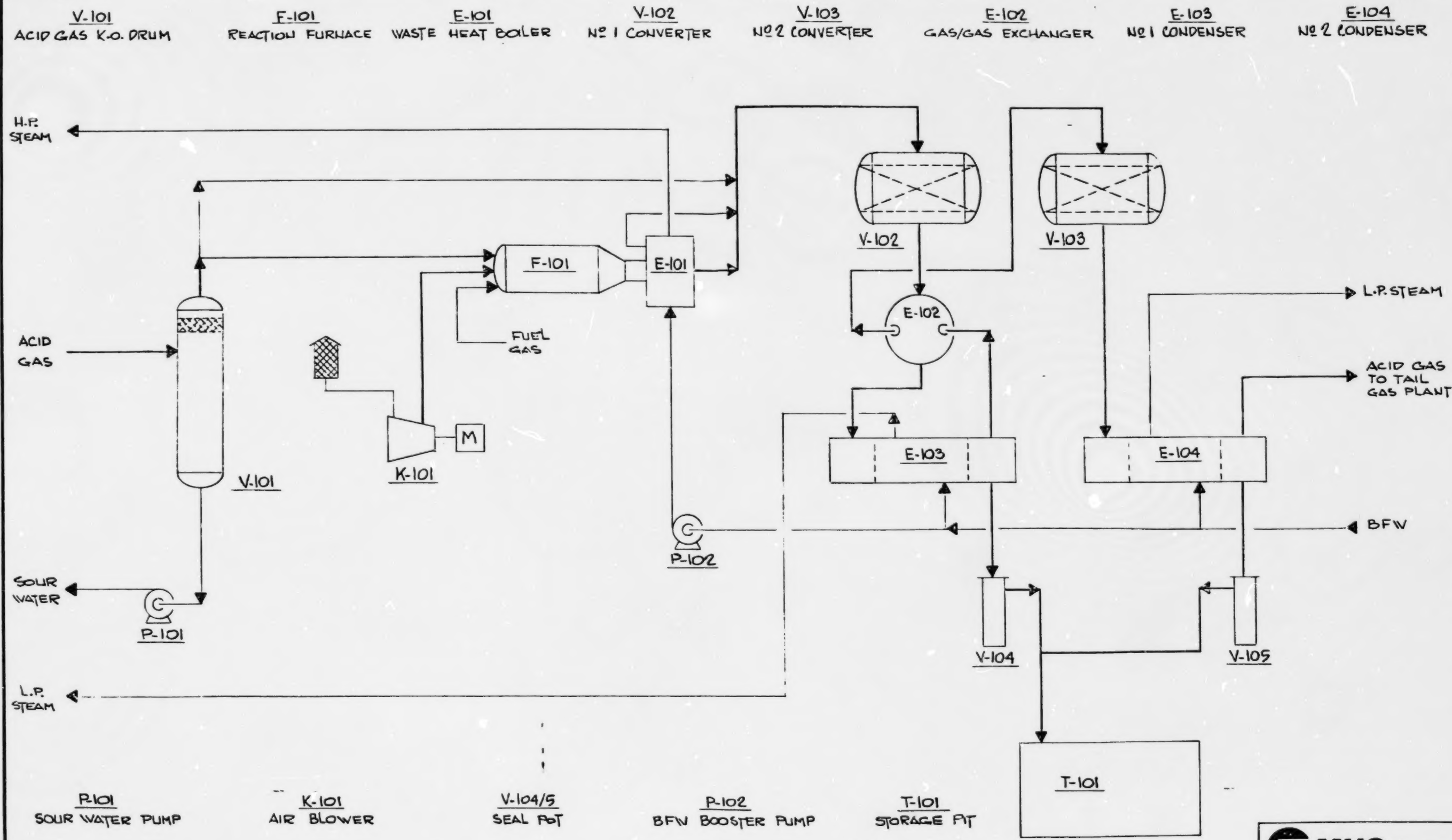
A Turbo Expander (KT-101) is provided for power recovery from the F-102 effluent. After KT-101, the effluent gases are emitted to the atmosphere through stack (U-102).



SCALE	NONE	BASIC CLAUSS - 2 STAGE STRAIGHT	DATE	FEB 1983
DRAWN	A.V.V.	THROUGH PROCESS FLOW DIAGRAM	W.O.	FED-0254
CHECKED		SULPHUR RECOVERY UNIT	DWG. NO.	FIGURE 4.1
PASSED				



SCALE	NONE	BASIC CLAUS. 3 STAGE STRAIGHT	DATE	FEB. 1993
DRAWN	A.V.V.	TROUGH PROCESS FLOW DIAGRAM	W.O. FED. 0254	REV
CHECKED		SULPHUR RECOVERY UNIT	DWG. NO.	
PASSED			FIGURE 4.2	



SCALE	NONE	BASIC CLAUS - SPLIT FLOW	DATE	FEB. 1983
DRAWN	A.M.V.	PROCESS FLOW DIAGRAM	W.O.	FED-0254
CHECKED		SULPHUR RECOVERY UNIT	DWG. NO.	FIGURE 4.3
PASSED				

5.0 TAIL GAS CLEAN-UP PROCESSES

5.1 Sulfreen Process

(Refer to Process Flow Diagram Figure 5.1).

The sulfreen process as shown in the process flow diagram is essentially an extension of the Claus process, except that H_2S and SO_2 are made to react at temperatures below the sulfur dew point of the reaction gas mixture.



There are usually three reactors, two of which are in adsorption service while one is being regenerated.

The tail gas from Claus unit passes through possible parallel reactors where H_2S and SO_2 are converted to elemental sulphur which is adsorbed on the catalyst, favoring the Claus conversion.

When the catalyst is fully loaded with sulphur, it has to be regenerated. During the regeneration, hot regeneration gas enters the reactor and passes downwards through the catalyst bed, stripping the sulphur and carrying it out to the sulphur condenser. The gas is cooled and the condensed sulphur is drained to the sulphur pit.

The regeneration gas then flows to the circulation blower, which drives it through the furnace to bring up the temperature. The hot gas enters the reactor in regeneration. Heating continues until most of the sulfur has been recovered.

5.0 TAIL GAS CLEAN-UP PROCESSES (cont'd)

5.2 CBA Process

(Refer to Process Flow Diagram Figure 5.2)

The Amoco Cold Bed Absorption (CBA) process is essentially, like Sulfreen, an extension of the Claus reaction but at a temperature below the sulphur dew point. However, this process uses the effluent gas from No. 1 Claus converter to regenerate the sulphur loaded catalyst.

Figure No. 5.2 shows a schematic flow diagram for the CBA tail gas unit. The unit features two CBA reactors: one is operating on the clean-up cycle while the other is being regenerated.

The tail gas from the Claus unit flows to CBA No. 1 Reactor which is on the clean-up mode and where residual H_2S and SO_2 are further converted to sulphur at a temperature below the sulphur dew point. The sulphur produced is adsorbed on the catalyst and the effluent gas from the reactor is routed directly to the incinerator. After a fixed time period, the catalyst bed will be loaded sulphur. The reactor will then be switched to regeneration.

Regeneration of the catalyst bed involves a heating step followed by a cooling step. Hot effluent gas from No. 1 Claus converter is used to heat up and purge the sulphur from the catalyst bed. The regeneration gas is then cooled in the CBA condenser where liquid sulphur is condensed and routed via a seal pot to the sulphur storage pit.

The effluent gas from CBA condenser flows to the regeneration gas blower. During the heating stage, the discharge from the regeneration gas blower is recycled back to the Claus unit.

5.0 TAIL GAS CLEAN-UP PROCESSES (cont'd)

5.2 CBA PROCESS (cont'd)

It is combined with the effluent stream from No. 2 Condenser, heated and fed to No. 2 Claus converter. During the cooling stage of the regeneration cycle, the discharge from the regeneration gas blower is further cooled by the regeneration gas cooler and circulated through the catalyst bed.

5.3 SCOT Process

(Refer to Schematic Flow Diagram, Figure 5.3).

This SCOT process consists of essentially three sections: heating and reduction, cooling and quenching, and H_2S absorption, stripping and recycle.

In the first section, the Claus tail gas is heated and reacted with a reducing gas over a fixed bed of Co-Mo catalyst in SCOT reactor. The reducing gas can be supplied from an outside source or generated by partial combustion of fuel gas in a SCOT inline burner. By this process virtually all sulphur compounds are reduced to H_2S .

In the second section, the effluent gas from the SCOT reactor is first cooled in an exchanger generating low pressure steam and then quenched in a quench column.

In the final section, the cooled reduced tail gas is counter-currently washed with a selective - alkanolamine solution DIPA or MDEA in an absorption tower where practically all H_2S is removed. The vent gas from SCOT absorber containing about 200 - 500 ppmv of H_2S is normally incinerated before discharging to the atmosphere.

5.0 TAIL GAS CLEAN-UP PROCESSES (cont'd)

5.3 SCOT Process (cont'd)

The rich amine solution from the bottom of the absorber is preheated by lean amine and fed to an amine stripper where the rich amine is regenerated. The recovered acid gas is recycled to the Claus unit.

5.4 BSR/MDEA Process

(Refer to Schematic Flow Diagram, Figure 5.4).

The BSR/MDEA process is very similar to the SCOT process, consisting of three major sections.

In the first section, (the BSR section) the tail gas from Claus unit is heated to reaction temperature in the reducing gas generator by mixing with the hot combustion products of fuel gas and air. The gas mixture is then fed to the catalytic reactor containing a Co-Mo catalyst where all sulphur compounds (SO_2 , Sx , COS , CS_2) are converted to H_2S . Partial combustion of fuel gas in the reducing gas generator provides hydrogen and carbon monoxide required for the conversion.

In the second section the reduced gas stream enters the reactor effluent cooler, and is further cooled in the contact condenser. Cooling in contact condenser is accomplished by evaporating water in the lower section and condensing and cooling in the upper section.

In the third section the cooled hydrogenated tail gas enters a MDEA absorber, where essentially all of the H_2S and some of the CO_2 are absorbed by the cool countercurrent MDEA solution. The residual gas, containing a small amount of H_2S , is normally incinerated before being discharged to the stack.

5.0 TAIL GAS CLEAN-UP PROCESSES (cont'd)

5.4 BSR/MDEA Process (cont'd)

The rich MDEA from the bottom of the absorber is preheated by lean MDEA and fed to the MDEA stripper, where MDEA solution is regenerated. The recovered acid gas recycled to the Claus unit.

5.5 BSR/Selectox Process

(Refer to Schematic Flow Diagram, Figure 5.5).

The BSR/Selectox process consists of a BSR section, a cooling section and a Selectox section.

The BSR section is identical in process to that described in 5.4. In the cooling section, the reduced gas stream enters the reactor effluent cooler, and is further cooled in the contact condenser by circulating water countercurrent to the gas. The purpose of cooling is to remove water thereby increasing the sulphur conversion efficiency in the next section.

In the Selectox section, the cooled hydrogenated tail gas is reheated, mixed with air, and fed to the Selectox reactor where H_2S is directly oxidized to elemental sulphur:



Some SO_2 may also be produced depending on the amount of air supplied. The gas is then cooled in the sulphur condenser where condensed sulphur is drained to the sulphur pit. The tail gas from sulphur condenser goes through an incinerator or waste heat boiler and to the stack.

5.0 TAIL GAS CLEAN-UP PROCESSES (cont'd)

5.6 BSR/Stretford Process

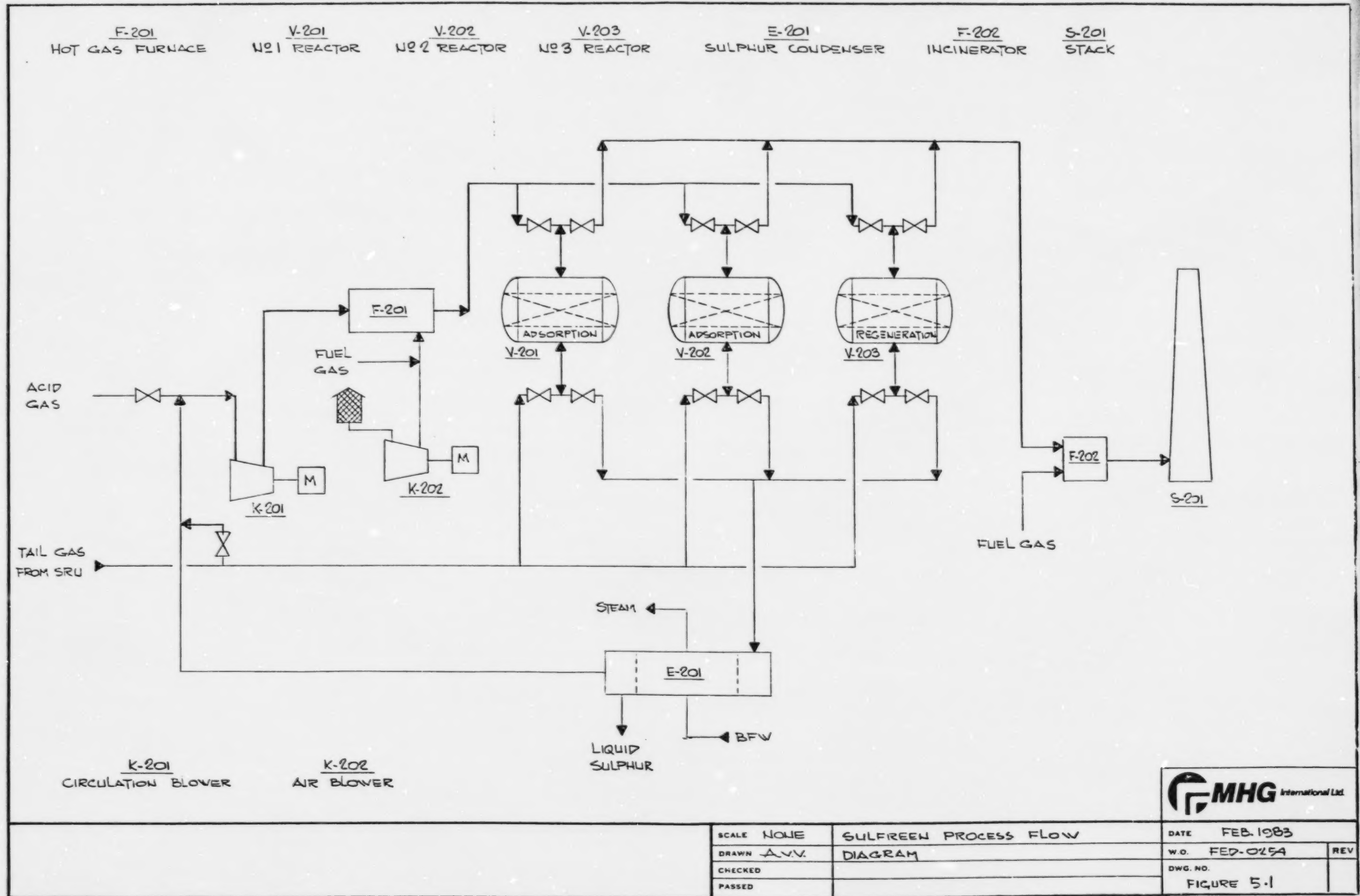
(Refer to Schematic Flow Diagram, Figure 5.6).

The BSR/Stretford process can also be described in three sections: a reduction section, a cooling section and Stretford section.

The reduction and cooling sections are identical to those of BSR/MDEA process described in Section 5.4.

In the Stretford section the cooled hydrogenated tail gas is fed to a venturi contactor and an absorber, where H_2S is removed almost quantitatively by the circulating Stretford solution. The process involves absorption of the H_2S in an oxidizing alkaline solution of sodium carbonate. The oxidizing agents in the solution, sodium vanadate, eventually convert H_2S to elemental sulphur. The solution is regenerated by contacting with air in the oxidizer tank where sulphur is floated off as a slurry. After separating the sulphur from the chemicals by filtering or centrifuging, with water washing, it is reslurried with wash water and heated to melt the sulphur. The molten sulphur flows from the decanter to the sulphur pit. The chemicals are returned to the system and the wash water is discarded.

The off-gas from the absorber, typically containing 20 - 80 ppmv carbonyl sulfide and less than 1 ppmv of H_2S , is normally vented to the atmosphere without incineration. A stand-by incinerator is normally provided for rare occasions when the H_2S content exceeds 10 ppm.



DATE	FEB. 1983	
W.O.	FED-025A	REV
DWG. NO.	FIGURE 5.1	

SCALE	NONE	SULFREEN PROCESS FLOW
DRAWN	AVV	DIAGRAM
CHECKED		
PASSED		

V-201
Nº1 CBA REACTOR

V-202
Nº2 CBA REACTOR

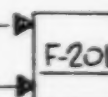
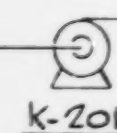
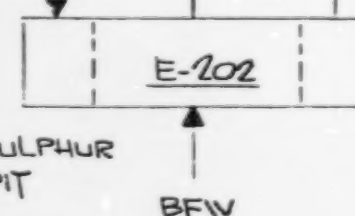
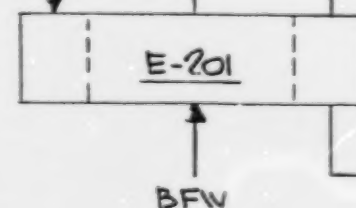
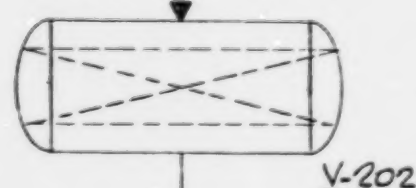
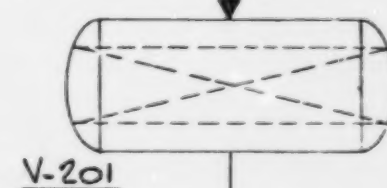
E-201
CBA CONDENSER

E-202
REGENERATOR
GAS COOLER

F-201
INCINERATOR

U-201
STACK

TAIL GAS
FROM
CLAUS UNIT
HOT EFFLUENT
FROM
Nº1 CLAUS CONVERTER



K-201
REGEN. GAS
BLOWER

FUEL GAS

REGEN. GAS TO
CLAUS UNIT

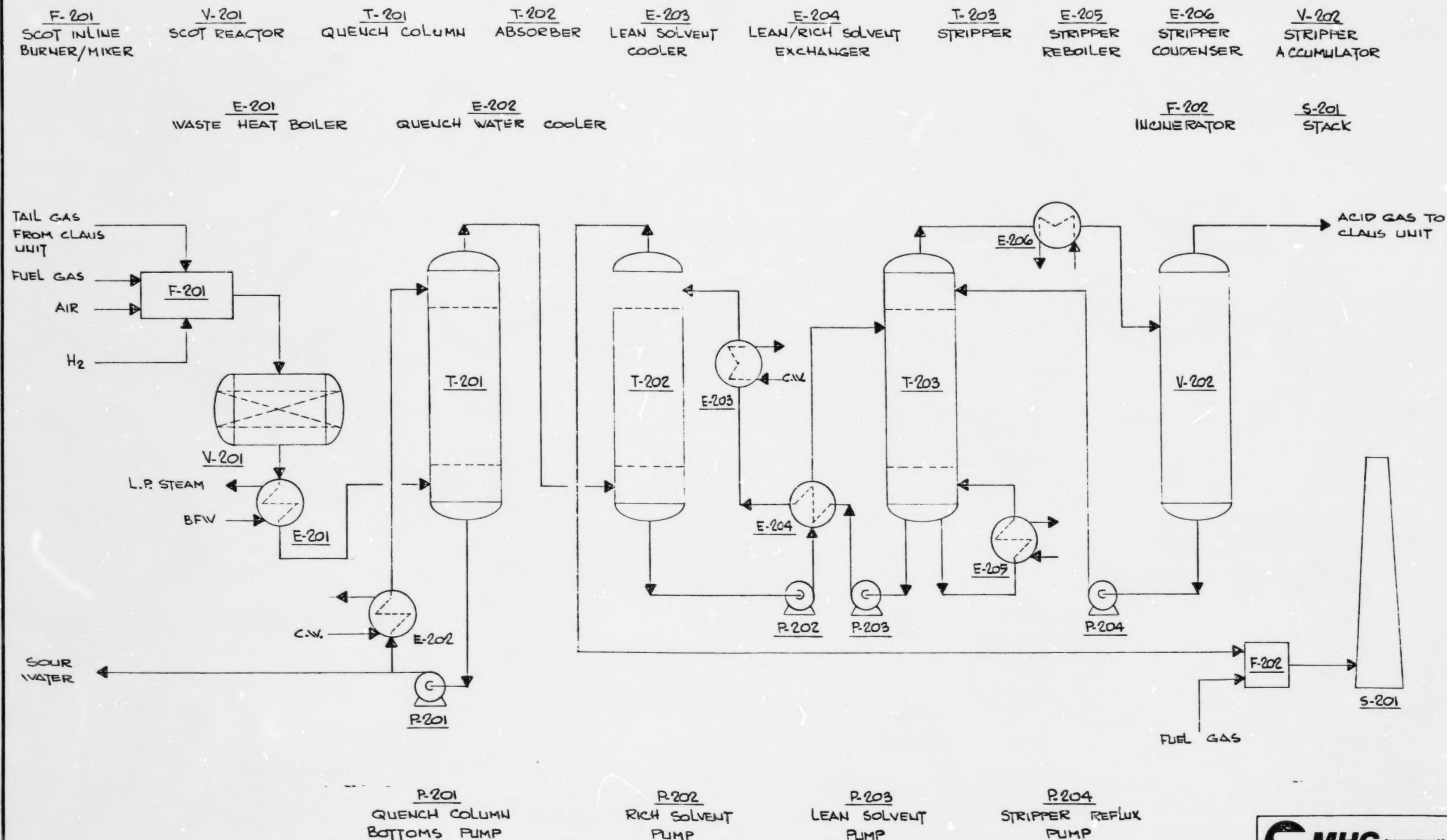
PREHEATED BFW
TO CLAUS UNIT

LIQ. SULPHUR
TO PIT



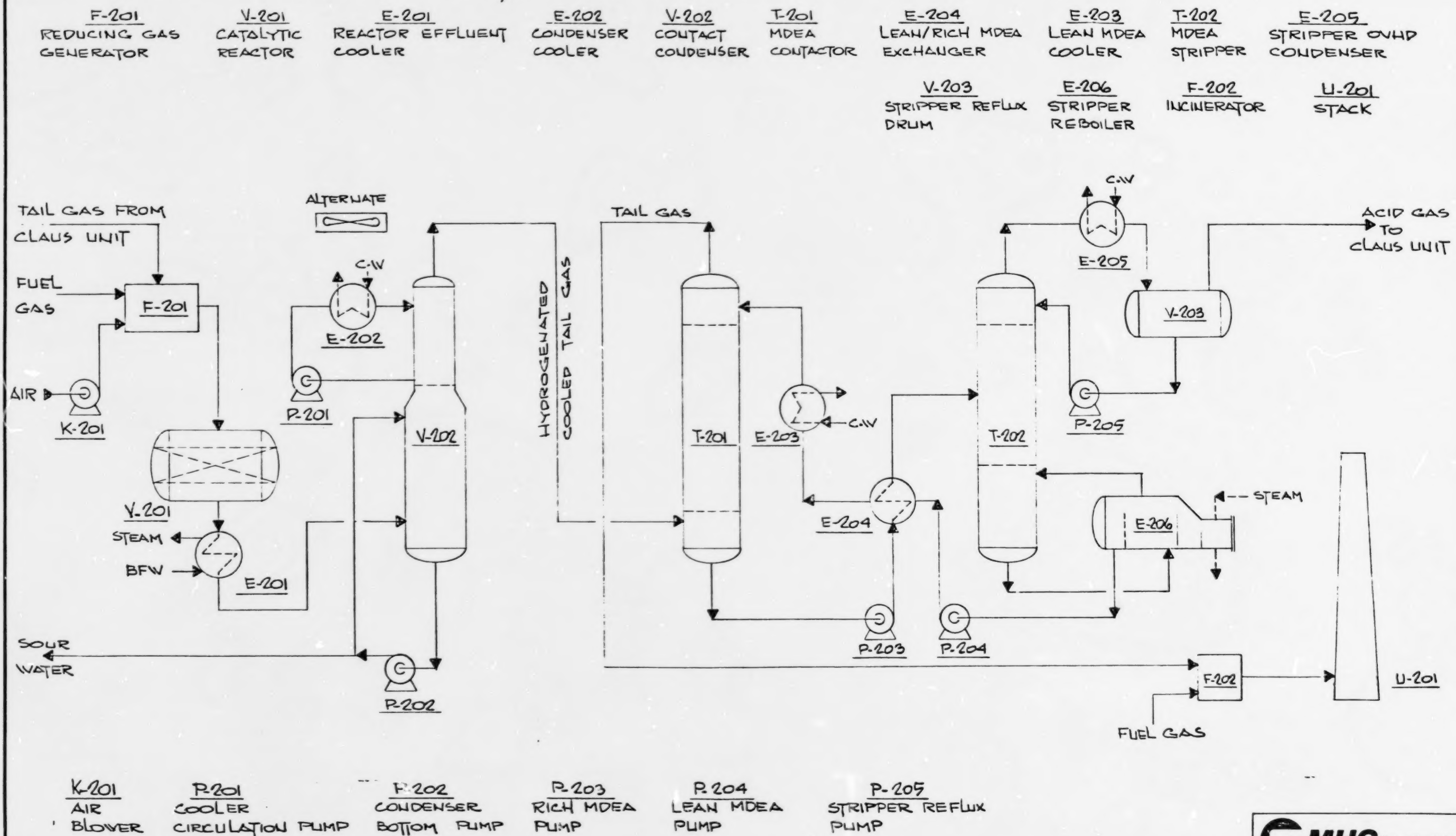
SCALE	NONE	CBA - PROCESS FLOW DIAGRAM
DRAWN	A.V.V.	
CHECKED		
PASSED		

DATE	MARCH 1983
W.O.	FED-0254
DWG. NO.	FIGURE 5-2
REV	



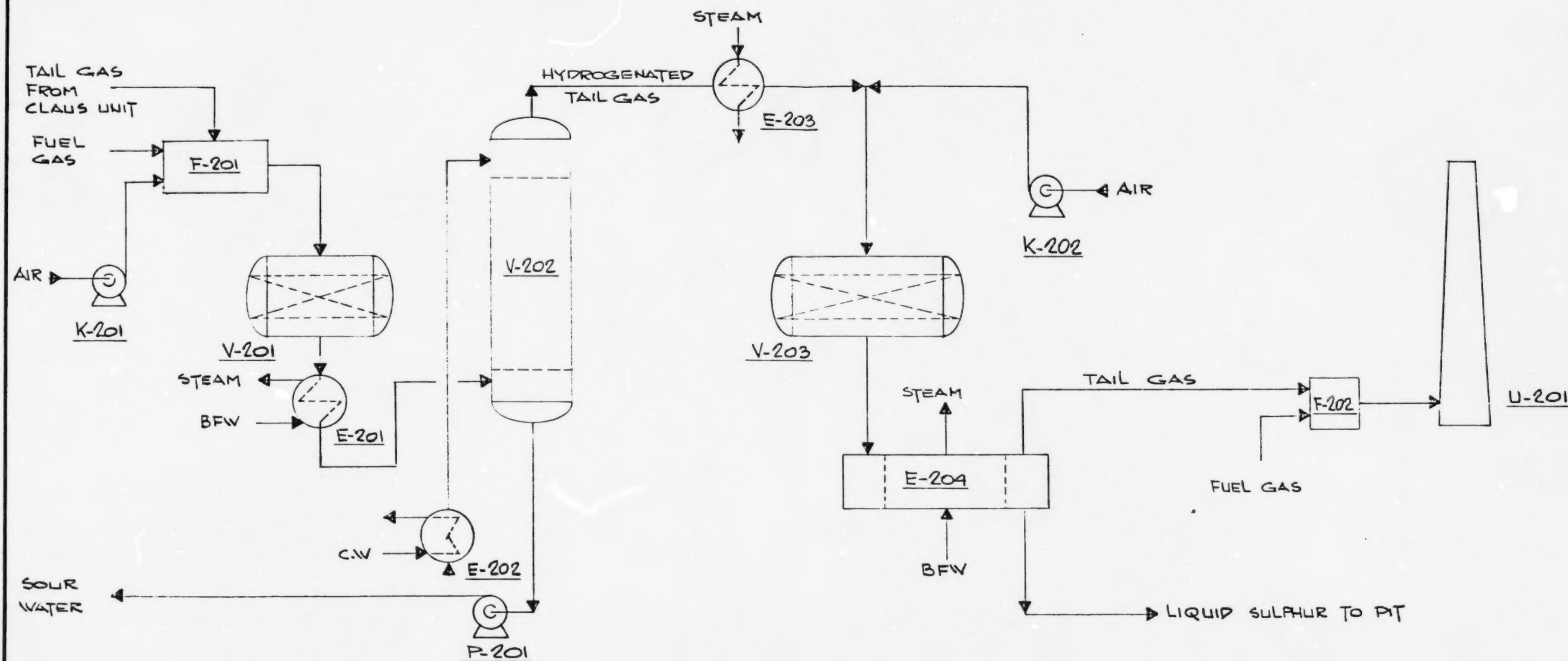
MHG International Ltd

SCALE	NONE	SCOT - PROCESS FLOW DIAGRAM	DATE	FEB. 1983
DRAWN	A.V.V.		W.O.	FED 0254
CHECKED			DWG. NO.	FIGURE 5-3
PASSED			REV	



SCALE	NONE	BSR/MDEA PROCESS FLOW DIAGRAM	DATE	MARCH 1983
DRAWN	A.V.V.		W.O.	FED-0254
CHECKED			DWG. NO.	FIGURE 5.4
PASSED				

<u>F-201</u> REDUCING GAS GENERATOR	<u>V-201</u> CATALYTIC REACTOR	<u>E-201</u> REACTOR EFFLUENT COOLER	<u>V-202</u> CONTACT CONDENSER	<u>E-202</u> CONDENSER COOLER	<u>E-203</u> REDUCED GAS REHEATER	<u>V-203</u> SELECTOX REACTOR	<u>E-204</u> SULPHUR CONDENSER	<u>F-202</u> INCINERATOR	<u>U-201</u> STACK
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K-201
AIR BLOWER

P-201
RECIRCULATION
WATER PUMP

K-202
AIR BLOWER



SCALE	NONE	BSR/SELECTOX PROCESS FLOW
DRAWN	A.V.V	DIAGRAM
CHECKED		
PASSED		

DATE	MARCH 1983
W.O.	FED-0254
DWG. NO.	FIGURE 5-5

6.0 TAIL GAS INCINERATION AND STACKS

Exit gas from tail gas clean-up unit contains H_2S , SO_2 , COS and CS_2 contaminants in it. Environmental regulations do not permit the emission of H_2S as such to atmosphere. Therefore, after meeting 97% or 98%+ sulphur recovery as required, the effluent is further incinerated with excess air and fuel gas. Burning of fuel gas with excess oxygen to raise the incinerator exit gas temperature to $538^{\circ}C$ is required to ensure all sulphur compounds are converted to SO_2 .

Incinerated gas is emitted to atmosphere through a stack. Two sizes of stacks 76m and 152m height are included. The required height of stack for a sulphur recovery plant depends on the location, surrounding and permissible ground level concentration.

In case of plant capacity less than 100 tonnes per day for 80/20 H_2S/CO_2 ratio, no tail gas unit is required. Therefore, tail gases from last condenser in Claus plant is directly incinerated and emitted to atmosphere through stack. Cost data for incinerator and stack for each case are provided in Figures.7-10 and 7-11.

7.0 CAPITAL COSTS

The Claus unit and sulphur plant installed costs, are shown in Figure 7-1 to Figure 7-9. The installed costs are based on August, 1982 Calgary price and includes all engineering, material and construction fees. The maximum single train capacity for different processes are shown on Table 7.1. When the sulphur input is higher than the capacity shown on Table 7.1, the multiple trains are used for the installed cost.

The installed costs for incinerator and stack are shown on Figure 7-10 and Figure 7-11.

The summary of installed costs of the selected 29 cases are shown on Table 7-2, 7-3 and 7-4.

TABLE 7-1
SINGLE TRAIN MAXIMUM CAPACITY OF CLAUS AND TAIL
GAS CLEAN-UP UNIT

H ₂ S/CO ₂ Ratio Acid Gas	Sulphur Input (tonnes/day)		
	15/85	50/50	80/20
<u>Sulphur Recovery Unit</u>			
Claus	400	1000	1250
<u>Tail Gas Processes</u>			
CBA	400	1000	1250
Sulfreen	800	2000	2500
SCOT	800	2000	2500
BSR/MDEA	800	2000	2500
BSR/Selectox	400	1000	1250
BSR/Stretford	400	1000	1250

TABLE 7-2
SUMMARY OF INSTALLED COSTS (M\$) ACID GAS H₂S/CO₂ RATIO 15/85

Acid Gas H ₂ S/CO ₂ Ratio Case No. Sulphur Input (tonnes/day)	15/85							
	3	4	5	6	7	8		
	150	150	150	150	150	150	150	150
<u>Sulphur Recovery Process</u>								
Claus, 3 Stages								
Claus, 2 Stages			11.2	11.2		9.9		11.2
	9.9	9.9						
<u>Tail Gas Processes</u>								
CBA	3.8							
Sulfreen		4.6						
SCOT			14.1					
BSR/MDEA				14.1				
BSR/Selectox					7.15			
BSR/Stretford						12.8		
<u>Incinerator & Stack</u>								
Incinerator	0.070	0.070	0.07	0.07	0.07	0.07	-	-
Stack, 152m	0.92	0.92	0.92	0.92	0.92	0.92	0.35*	0.35*
TOTAL INVESTMENT	14.69	15.49	26.29	26.29	18.04	24.35		

* In case No. 8 stack height of 30m is considered.

TABLE 7-3

SUMMARY OF INSTALLED COSTS (M\$) ACID GAS H₂S/CO₂ RATIO 50/50

Acid Gas H ₂ S/CO ₂ Ratio Case No.	50/50					
	11	12	13	14	15	16
Sulphur Input (tonnes/day)	350	350	350	350	350	350
<u>Sulphur Recovery Process</u>						
Claus, 3 Stages						
Claus, 2 Stages			14.5	14.5		14.5
	12.7	12.7			12.7	
<u>Tail Gas Processes</u>						
CBA	3.8					
Sulfreen		4.98				
SCOT			14.4			
BSR/MDEA				14.4		
BSR/Selectox					7.25	
BSR/Stretford						12.9
<u>Incinerator & Stack</u>						
Incinerator	0.07	0.07	0.07	0.07	0.07	-
Stack, 152m	0.84	0.84	0.84	0.84	0.84	0.32*
TOTAL INVESTMENT	17.41	18.59	29.81	29.81	20.86	27.72

* In case No. 16 stack height of 30m is considered.

TABLE 7-4

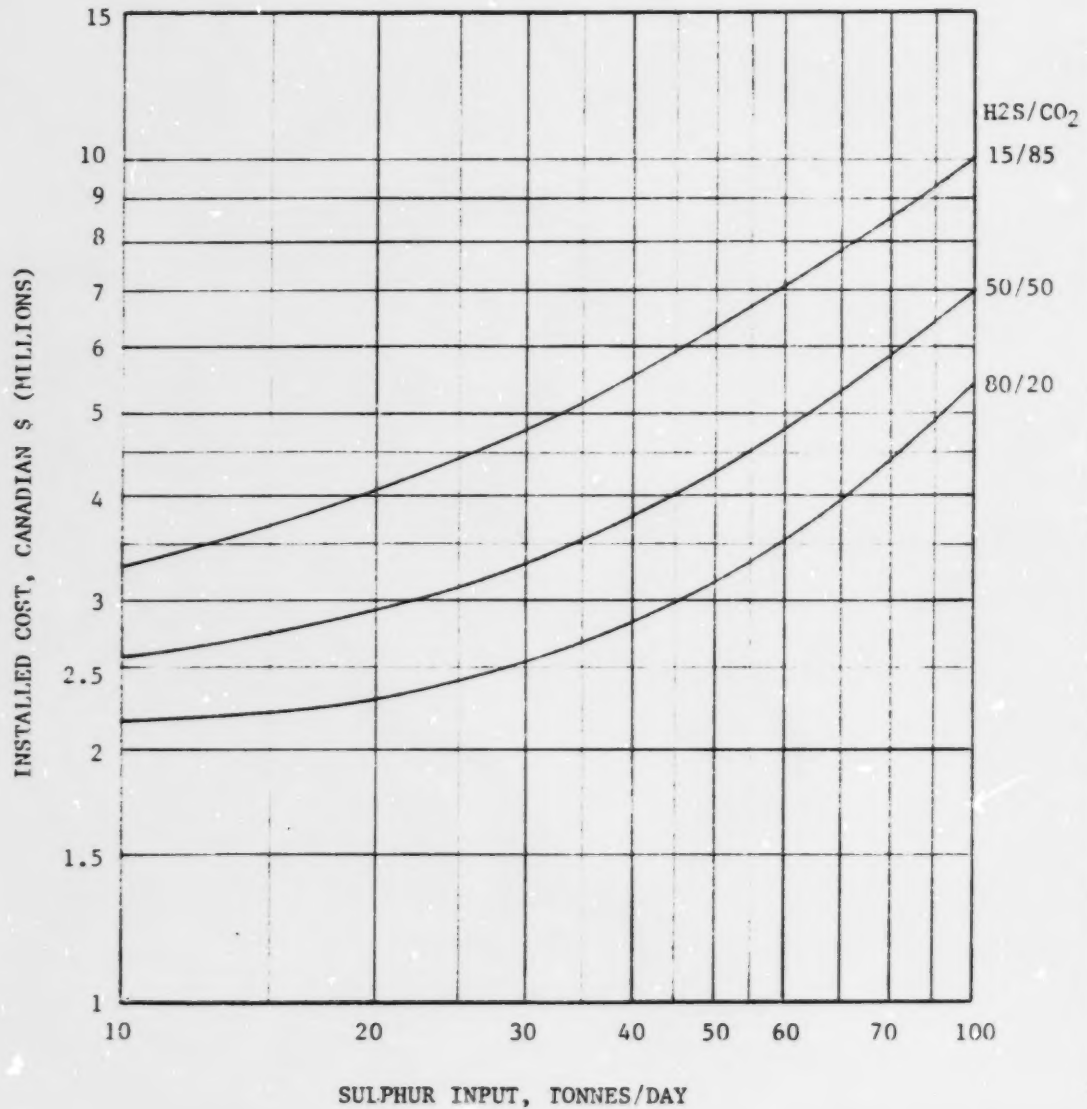
SUMMARY OF INSTALLED COSTS (M\$) ACID GAS H₂S/CO₂ RATIO 80/20

Acid Gas H ₂ S/CO ₂ Ratio Case No. Sulphur Input tonnes/day	80/20												
	17 400	18 400	19 400	20 400	21 400	22 400	23 400	24 1000	25 1000	26 1000	27 1000	28 1000	29 1000
<u>Sulphur Recovery Process</u>													
Claus, 3 Stages	5.4												
Claus, 2 Stages		12.6	12.6	14.0	14.0	12.6	14.0	23.8	23.8	26.4	26.4	23.8	26.4
<u>Tail Gas Processes</u>													
CBA		3.8						6.77	8.2				
Sulfreen			4.90							24.1			
SCOT				13.8							24.1		
BSR/MDEA					13.8							12.0	18.4
BSR/Selectox						6.70							
BSR/Stretford							12.6						
<u>Incinerator & Stack</u>													
Incinerator	0.04	0.07	0.07	0.07	0.07	0.07	-	0.10	0.10	0.10	0.10	0.10	-
Stack, 152m	0.76	0.76	0.76	0.76	0.76	0.76	0.29*	1.14	1.14	1.14	1.14	1.14	.43*
Total Investment	6.2	17.23	18.3	28.63	28.63	20.33	26.89	31.81	33.24	51.74	51.74	37.04	45.23

* In case No. 23 and 29 stack height of 30m is considered.

FIG. 7.1

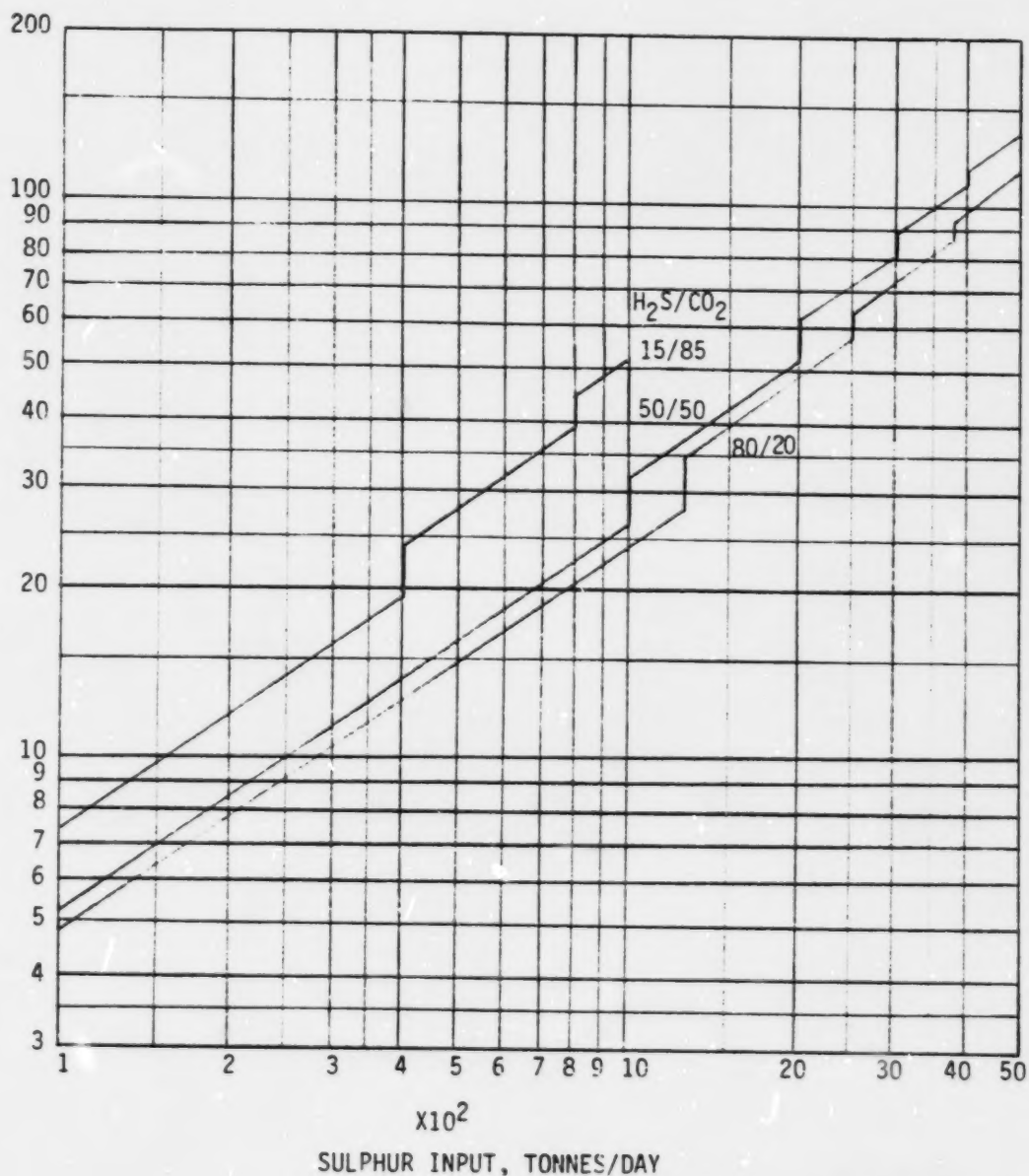
SULPHUR PLANT INSTALLED COST 10-100 TONNES/DAY
15%, 50% & 80% H₂S, 97% RECOVERY



SCALE		DATE	MARCH, 1983	REV.
DRAWN		W.O.	FED-0254	
CHECKED		DWG. NO.		
PASSED				

FIGURE 7.2

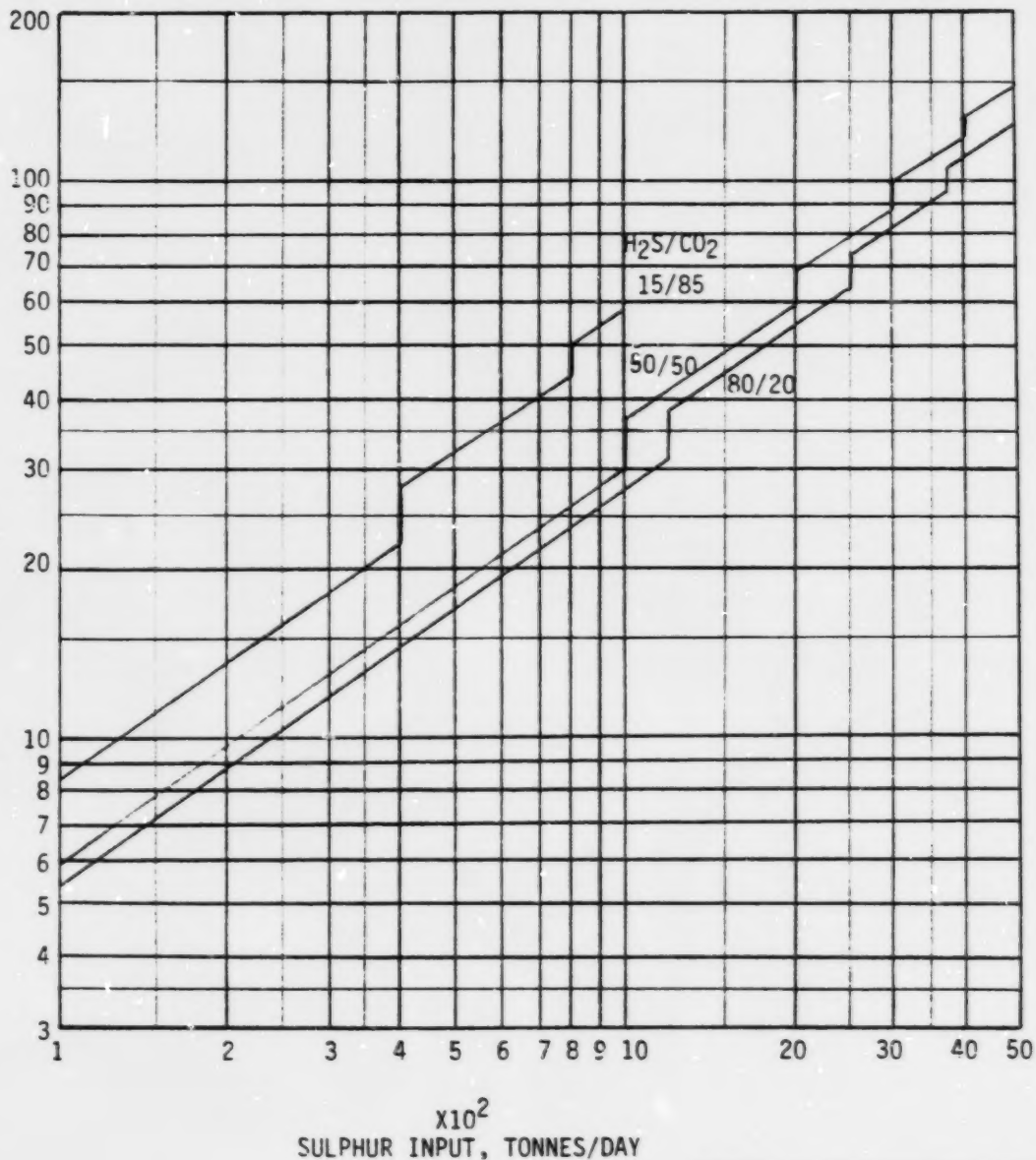
TWO STAGE
(CLAUS) SULPHUR PLANT INSTALLED COST. 100-5000 TONNES/DAY
15%, 50% & 80% H₂S IN FEED GAS



SCALE		DATE	
DRAWN		W.O.	REV.
CHECKED		DWG. NO.	
PASSED			

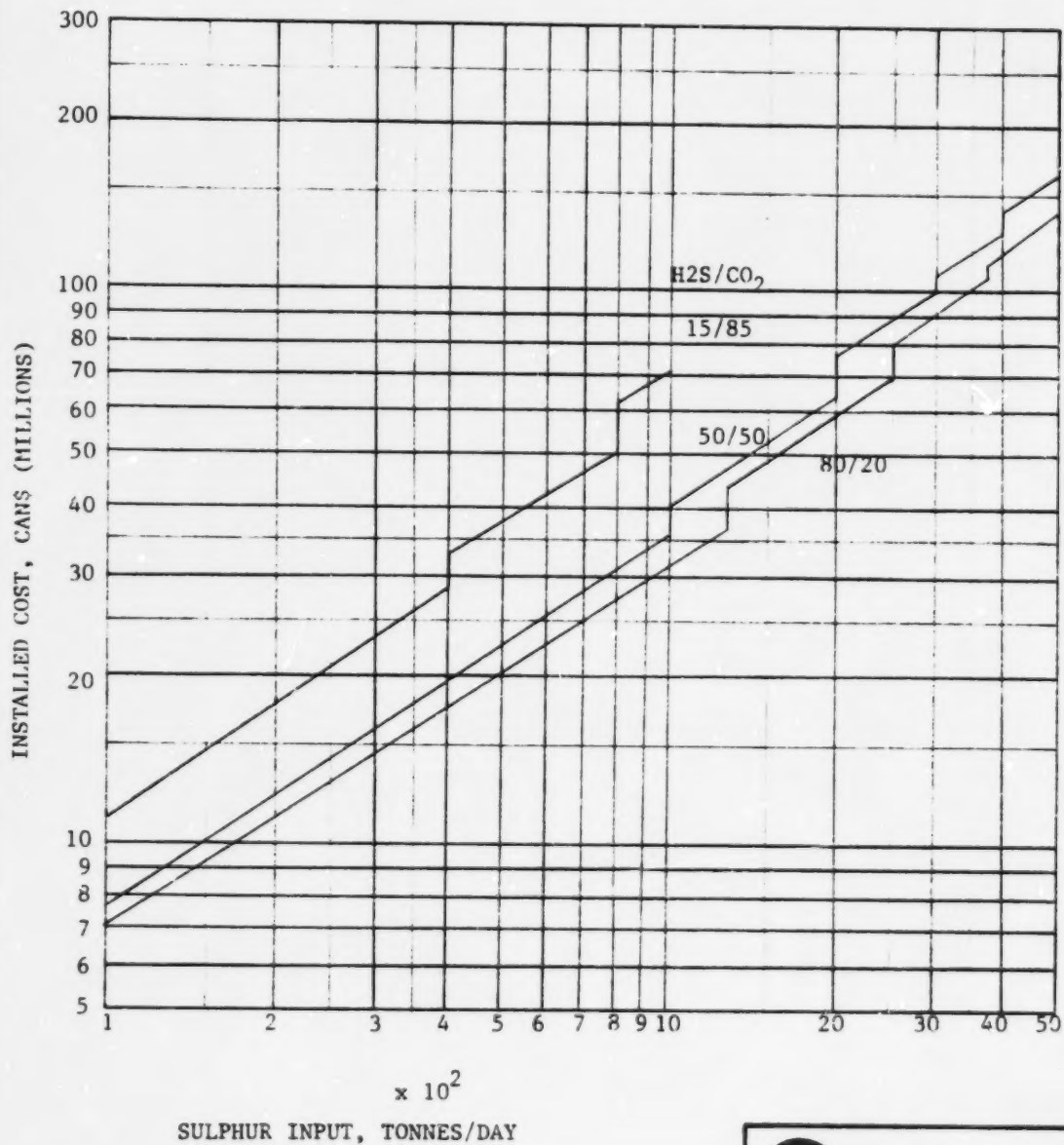
FIGURE 7.3

THREE STAGE
(CLAUS) SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50% & 80% H₂S IN FEED GAS



SCALE		DATE	
DRAWN		W.O.	REV.
CHECKED		DWG. NO.	
PASSED			

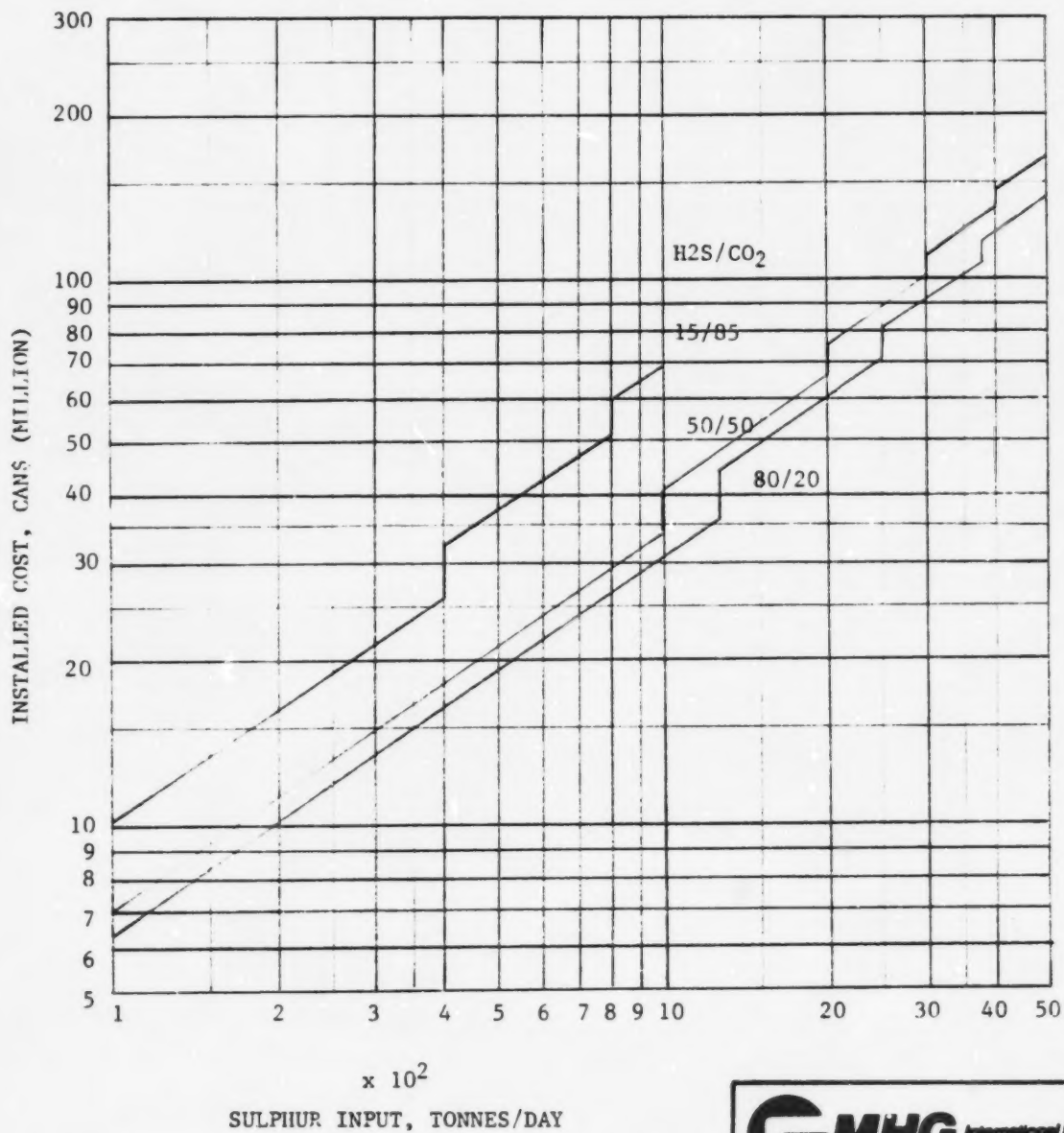
FIG. 7.4
SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50% & 80% H₂S, CLAUS + SULFREEN



SCALE		DATE	MARCH, 1983	
DRAWN		W.D.	FED-0254	REV.
CHECKED		DWS. NO.		
PASSED				

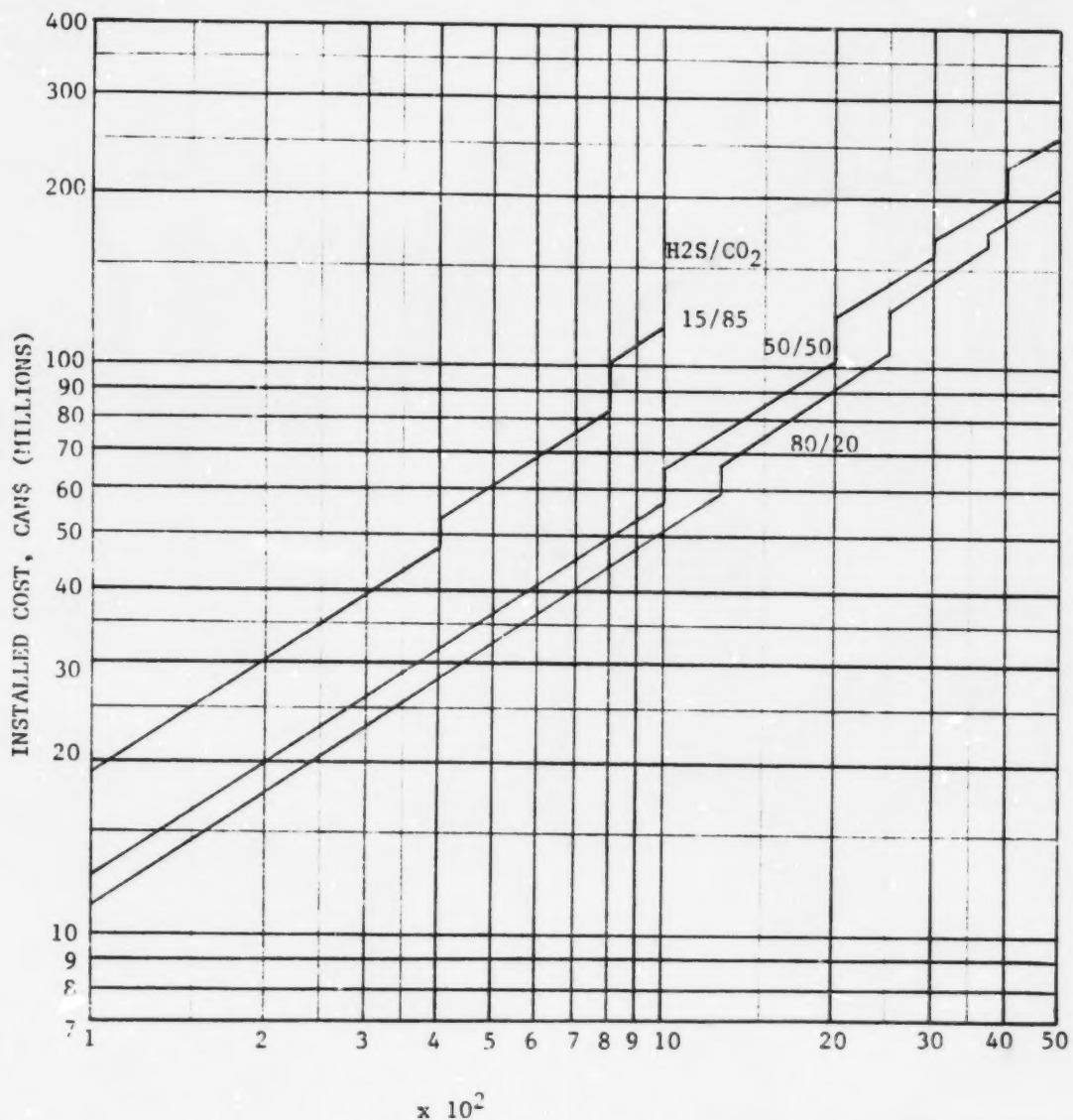
FIG. 7.5

SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50%, & 80% H₂S CLAUS + CBA



SCALE		DATE	MARCH, 1983	
DRAWN		W.D.	FED-0254	REV.
CHECKED		DWG. NO.		
PASSED				

FIG. 7.6
SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50% & 80% H₂S CLAUS + SCOT



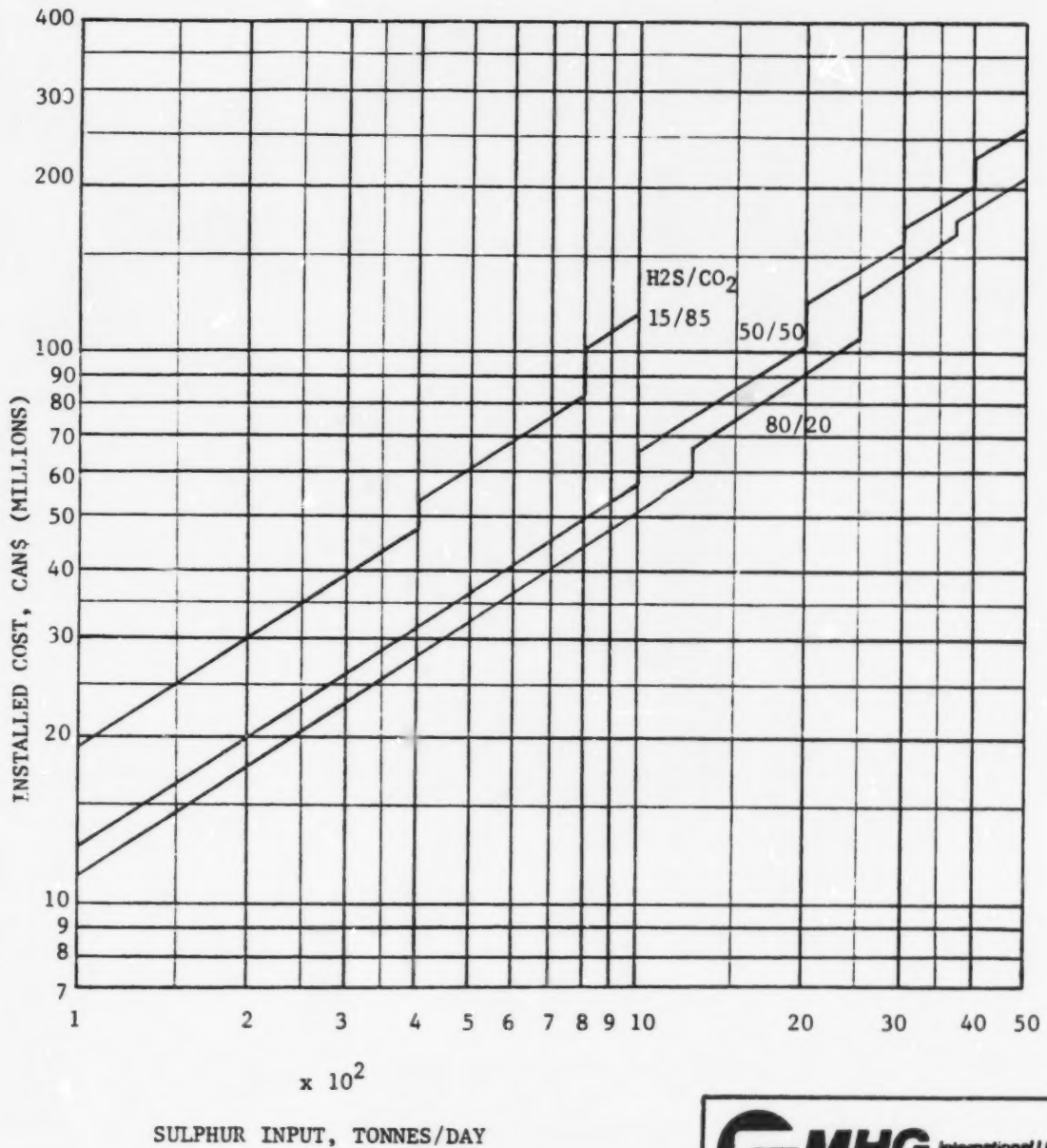
SULPHUR INPUT TONNES/DAY



SCALE		DATE	MARCH, 1983	
DRAWN		W.O.	FED-0254	REV.
CHECKED		DWS. NO.		
PASSED				

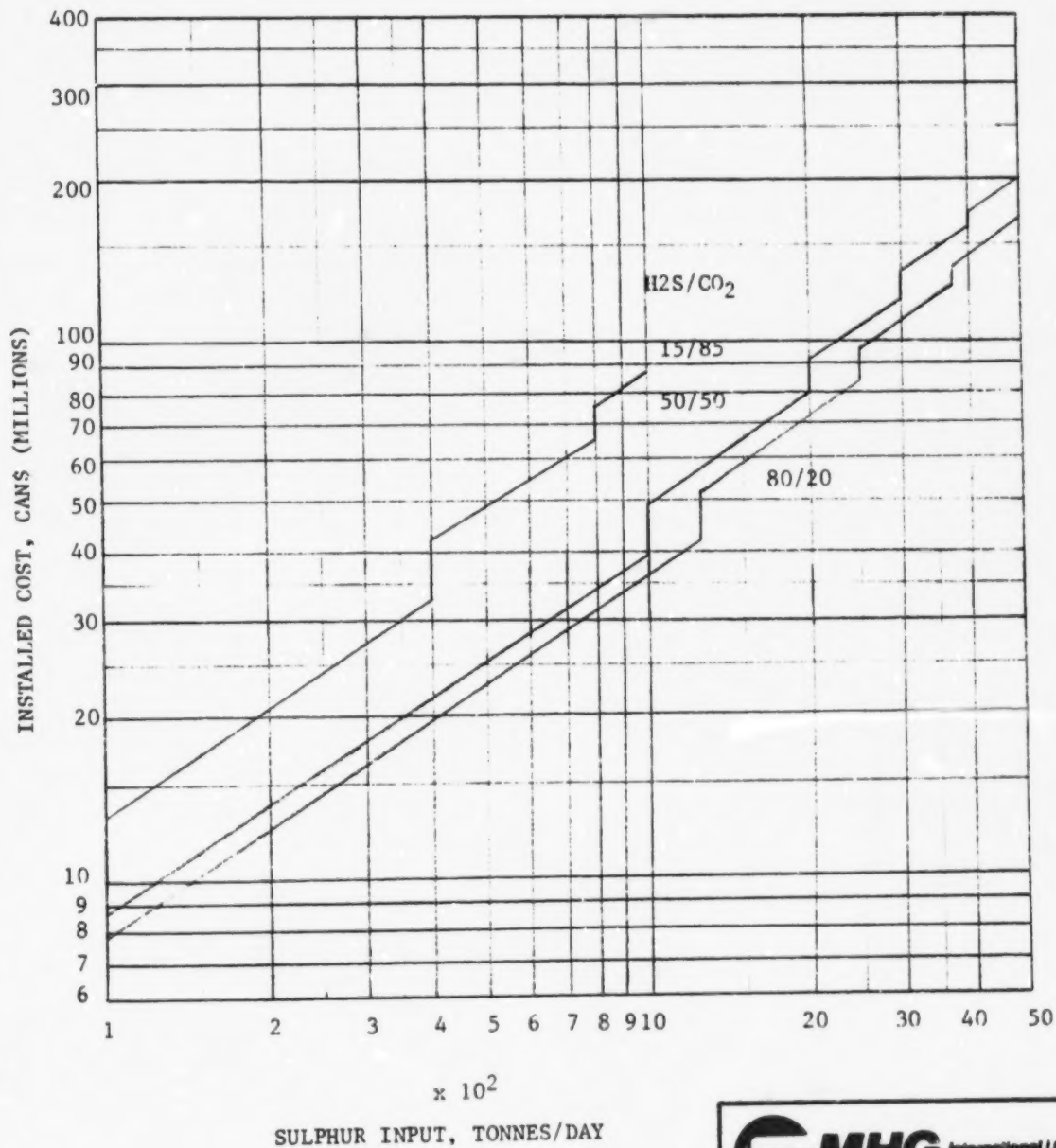
FIG. 7.7

SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50%, 28% H₂S CLAUS + BSR/MDEA



SCALE		DATE	MARCH, 1983
DRAWN		W.O.	FED-0254
CHECKED		DWG. NO.	
PASSED		REV.	

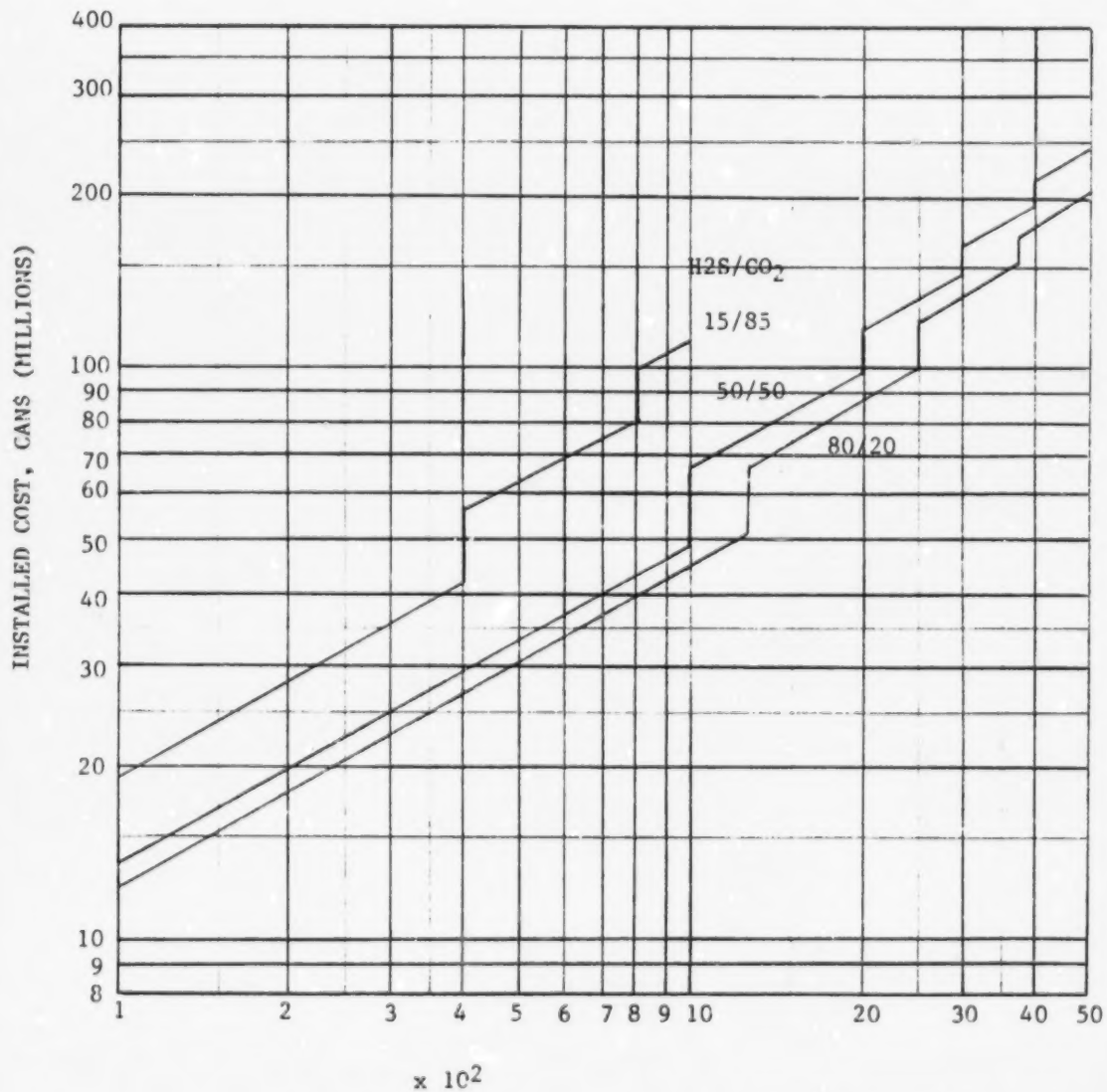
FIG. 7.8
SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50% & 80% H₂S, CLAUS + BSR/SELECTOX



SCALE		DATE	
DRAWN		W.O.	REV.
CHECKED		DWG. NO.	
PASSED			

FIG. 7.9

SULPHUR PLANT INSTALLED COST, 100-5000 TONNES/DAY
15%, 50%, & 85% H₂S, CLAUS + BSR/STRETTFORD



SULPHUR INPUT, TONNES/DAY



SCALE		DATE	MARCH, 1983	
DRAWN		W.O.	FED-0254	REV.
CHECKED		DWG. NO.		
PASSED				

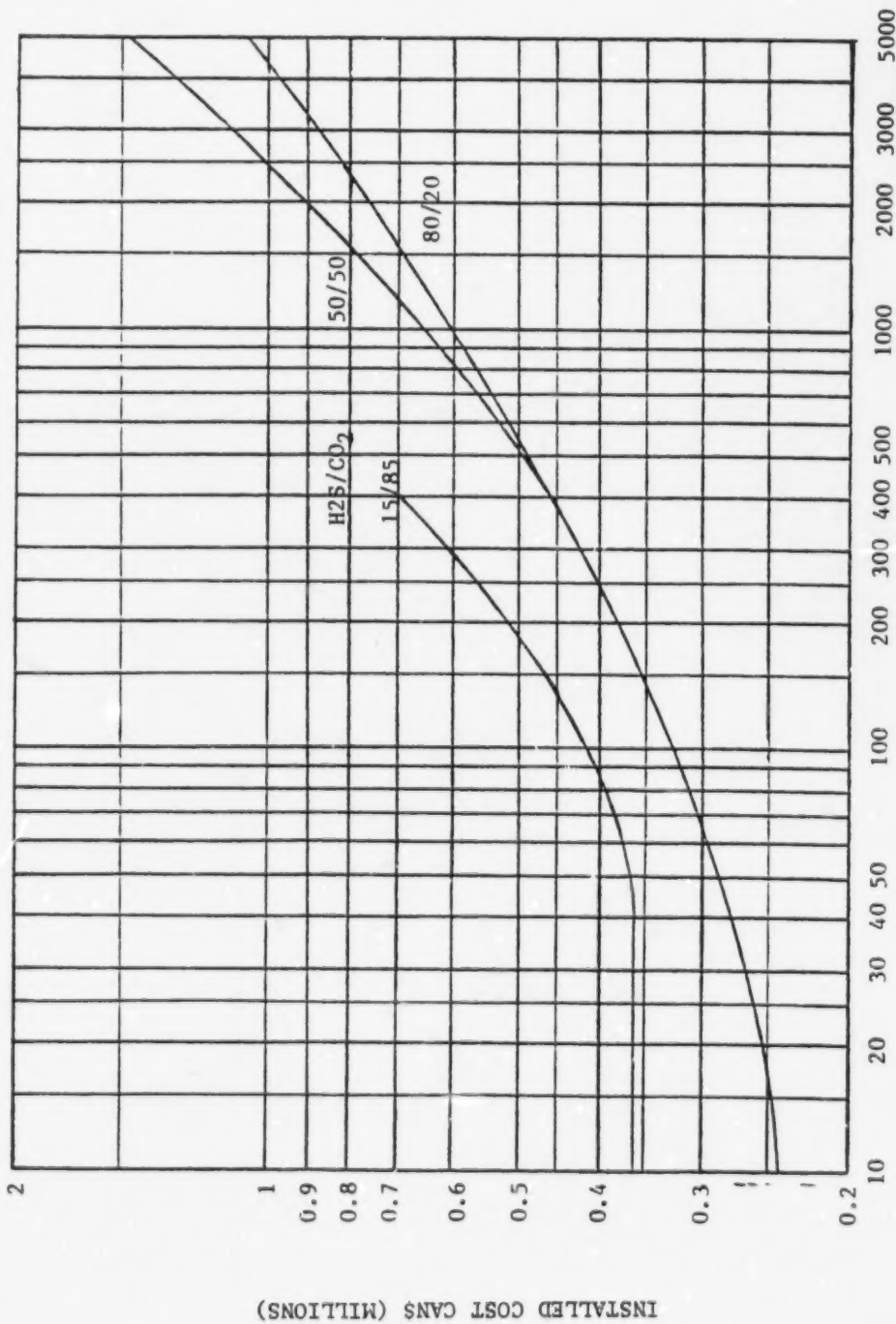
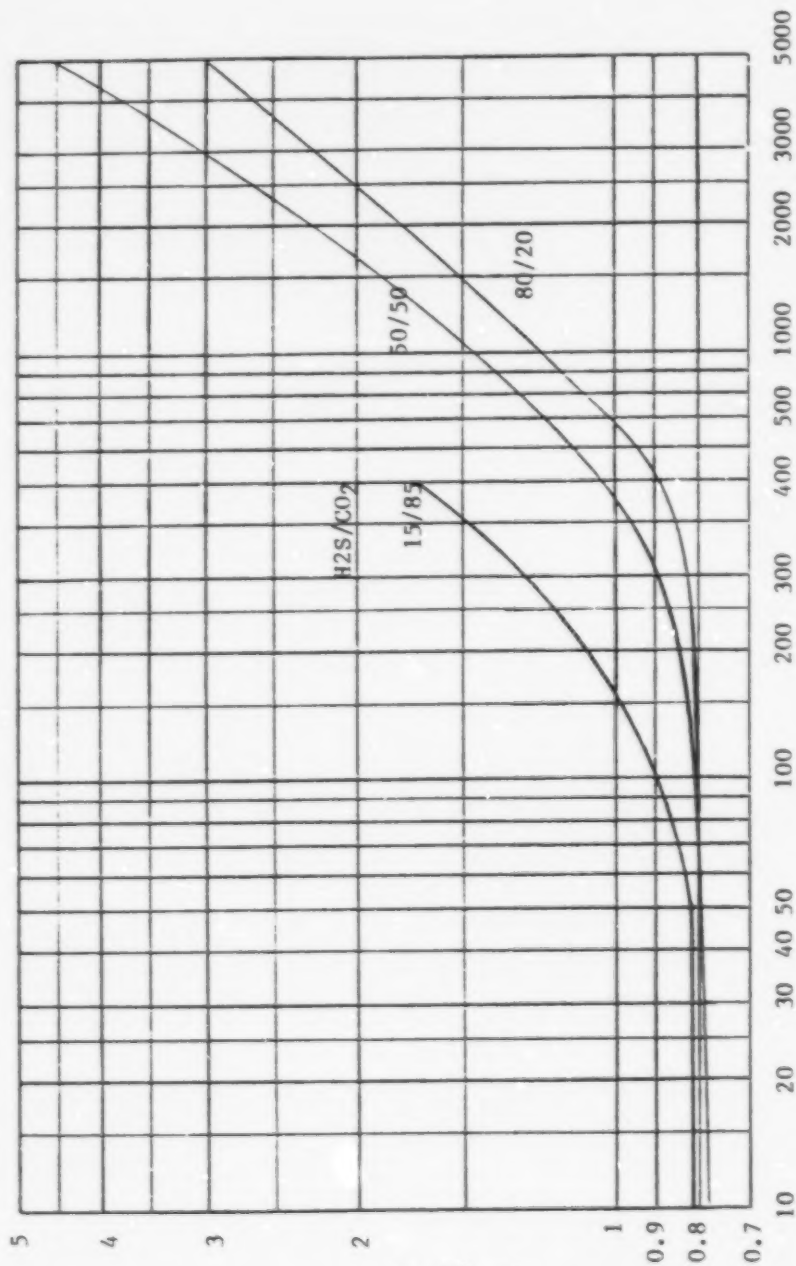


FIG. 7.10
INCINERATOR & STACK INSTALLED COST
15%, 50% & 80% H₂S, 76m High

MHG International Ltd

SCALE		DATE	MARCH, 1983
DRAWN		W.O.	FED-0254
CHECKED		DWG. NO.	
PASSED			



CMHG International Ltd.

8.0 OPERATING COSTS

The annual direct operating cost includes the cost for utilities, chemicals and catalyst, maintenance and repair and operating personnel. Table 8-1 presents the economic assumptions used to determine the annual direct operating cost of various cases. Fixed charges and plant overhead allowances are not included in that table. Table 8-2 shows the estimated utilities requirement and production summary along with the catalyst and chemical cost for the selected twenty nine cases. The annual direct operating cost and the product credit summary are listed in Table 8-3.

Table 8-1

ECONOMIC ASSUMPTIONS FOR ANNUAL DIRECT OPERATING COST ESTIMATION

A. Utility Prices

1. 4140 kPa(g) Saturated Steam	\$16.80/1000kg
2. 345 kPa(g) Saturated Steam	\$13.76/1000kg
3. 105 kPa(g) Saturated Steam	\$11.90/1000kg
4. Cooling Water	\$16.21/1000m ³
5. Steam Condensate	\$ 3.08/1000kg
6. Fuel Gas	\$ 3.88/10 ⁹ J
7. Power	\$ 0.05/KWH
8. Sulphur	\$70/t
9. Boiler Feed Water	\$ 3.50/1000kg

B. Chemicals and Catalyst

1. Alumina	\$1200/1000kg
2. Cobalt - Molybdenum	\$6000/m ³ or \$7.1/kg
3. MDEA	\$ 2.90/kg

C. Days of Operation in One Year

350 days

D. Maintenance and Repair

Labor and Material

3% of fixed capital

E. Operating Personnel

1. Operators	\$15/h
2/3 per shift for Claus Sulphur Recovery Unit. 10 t/d to 500 t/d.	
1 per shift for Claus sulphur Recovery Unit 500 t/d to 1000 t/d.	
2/3 per shift for Tail Gas Clean-up Unit.	
2. Supervisor	\$20/h
1/4 per shift for Sulphur Recovery Unit and Tail Gas Clean-up Unit.	

Table 8-2

UTILITIES AND CATALYST AND CHEMICAL SUMMARY

Case No.	1	2	3	4	5
Claus-Flow Scheme	Split Flow	Split Flow	Split Flow	Split Flow	Split Flow
-No. of Stgs.	2	2	2	2	3
Sulfur Input, t/d	50	50	150	150	150
H ₂ S/CO ₂ Ratio	15/85	15/85	15/85	15/85	15/85
TGCU Process Type	CBA	Sulfreen	CBA	Sulfreen	SCOT
Overall S-Rec. %	98.0	98.0	98.0	98.0	99.9
<u>Utilities Required</u>					
Treated BFW, kg/h	6,380	6,380	19,140	19,140	21,490
Electric Power, KW	182	182	546	546	430
Fuel Gas, m ³ (st)/h	399	399	1,197	1,197	1,600
Steam, 345 kPa(ga) kg/h*	-	-	-	-	570
Cooling Water, m ³ /h	-	-	-	-	870
<u>Utilities Produced</u>					
4140 kPa(ga) steam, kg/h	3,518	3,518	10,550	10,550	10,550
345 kPa(ga) steam, kg/h	1,262	1,262	3,790	3,790	-
105 kPa(ga) steam, kg/h	1,600	1,600	4,800	4,800	2,490
Steam Condensate, kg/h	-	-	-	-	-
<u>Catalyst and Chemical Costs</u>					
Combined Cost, \$/d	59	74	180	230	340

* Net consumption

Table 8-2

(cont'd)

UTILITIES AND CATALYST AND CHEMICAL SUMMARY

Case No.	6	7	8	9	10
Claus-Flow Scheme	Split Flow	Split Flow	Split Flow	Strgt.Thru	Strgt.Thru
-No.of Stgs.	3	3	2	2	2
Sulfur Input, t/d	150	150	150	15	15
H ₂ S/CO ₂ Ratio	15/85	15/85	15/85	50/50	50/50
TGCU Process Type	BSR/MDEA	BSR/ Selectox	BSR/ Stretford	CBA	Sulfreen
Overall S-Rec. %	99.9	97.8	99.98	98.5	98.5
Utilities Required					
Treated BFW, kg/h	21,490	22,710	21,830	1,750	1,750
Electric Power, KW	445	558	490	45	45
Fuel Gas, m ³ (st)/h	1,600	1,780	1,058	44.6	44.6
Steam, 345 kPa(ga) kg/h*	570	-	-	-	-
Cooling Water, m ³ /h	870	145	-	-	-
Utilities Produced					
4140 kPa(ga)steam, kg/h	10,550	10,550	10,550	1,076	1,076
345 kPa(ga)steam, kg/h	-	8,790	8,790	525	525
105 kPa(ga)steam, kg/h	2,490	3,370	2,490	99	99
Steam Condensate, kg/h	9,020	-	-	50	50
Catalyst and Chemical Costs					
Combined Cost, \$/d	229	300	678.5	9	11

* Net consumption.

Table 8-2

(cont'd)

UTILITIES AND CATALYST AND CHEMICAL SUMMARY

Case No.	11	12	13	14	15
Claus-Flow Scheme	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru
-No.of Stgs.	2	2	3	3	2
Sulfur Input, t/d	350	350	350	350	350
H ₂ S/CO ₂ Ratio	50/50	50/50	50/50	50/50	50/50
TGCU Process Type	CBA	Sulfreen	SCOT	BSR/MDEA	BSR/ Selectox
Overall S-Rec. %	98.5	98.5	99.9	99.9	98.5
Utilities Required					
Treated BFW, kg/h	40,820	40,820	43,150	43,150	44,400
Electric Power, KW	1,037	1,037	970	990	1,048
Fuel Gas, m ³ (st)/h	1,041	1,041	1,400	1,400	1,560
Steam, 345 kPa(ga) kg/h*	-	-	-	-	-
Cooling Water, m ³ /h	-	-	1,010	1,010	145
Utilities Produced					
4140 kPa(ga) steam, kg/h	25,100	25,100	25,100	25,100	25,100
345 kPa(ga) steam, kg/h	12,240	12,240	3,110	3,110	17,250
105 kPa(ga) steam, kg/h	2,310	2,310	-	-	875
Steam Condensate, kg/h	1,170	1,170	14,940	14,940	1,175
<u>Catalyst and Chemical Costs</u>					
Combined Cost, \$/d	186	228	375	269	300

* Net consumption.

Table 8-2

(cont'd)

UTILITIES AND CATALYST AND CHEMICAL SUMMARY

Case No.	16	17	18	19	20
Claus-Flow Scheme	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru
-No.of Stgs.	3	3	2	2	3
Sulfur Input, t/d	350	100	400	400	400
H ₂ S/CO ₂ Ratio	50/50	80/20	80/20	80/20	80/20
TGCU Process Type	BSR/ Stretford	Not Required	CBA	Sulfreen	SCOT
Overall S-Rec. %	99.98	97.0	98.7	98.7	99.9
Utilities Required					
Treated BFW, kg/h	43,530	14,180	59,220	59,220	61,380
Electric Power, KW	980	226	1,117	1,117	1,110
Fuel Gas, m ³ (st)/h	920	103	586	586	850
Steam, 345 kPa(ga) kg/h*	-	-	-	-	9,000
Cooling Water, m ³ /h	-	-	-	-	1,350
Utilities Produced					
4140 kPa(ga) steam, kg/h	25,100	10,510	45,680	45,680	45,680
345 kPa(ga) steam, kg/h	17,250	2,890	9,525	9,525	-
105 kPa(ga) steam, kg/h	-	-	2,310	2,310	-
Steam Condensate, kg/h	1,180	780	1,705	1,705	24,700
Catalyst and Chemical Costs					
Combined Cost, \$/d	678.5	33	186	228	460

* Net consumption.

Table 8-2

(cont'd)

UTILITIES AND CATALYST AND CHEMICAL SUMMARY

Case No.	21	22	23	24	25
Claus-Flow Scheme	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru
-No.of Stgs.	3	2	3	2	2
Sulfur Input, t/d	400	400	400	1000	1000
H ₂ S/CO ₂ Ratio	80/20	80/20	80/20	80/20	80/20
TGCU Process Type	BSR/MDEA	BSR/ Selectox	BSR/ Stretford	CBA	Sulfreen
Overall S-Rec. %	99.9	98.6	99.98	98.7	98.7
Utilities Required					
Treated BFW, kg/h	61,380	62,790	61,910	148,050	148,050
Electric Power, KW	-	1,128	1,060	2,563	2,563
Fuel Gas, m ³ (st)/h	850	879	518	1,465	1,465
Steam, 345 kPa(ga) kg/h*	9,000	-	-	-	-
Cooling Water, m ³ /h	1,350	145	-	-	-
Utilities Produced					
4140 kPa(ga)steam, kg/h	45,680	45,680	45,680	114,200	114,200
345 kPa(ga)steam, kg/h	-	14,530	14,530	23,810	23,810
105 kPa(ga)steam, kg/h	-	875	-	5,775	5,775
Steam Condensate, kg/h	24,700	1,705	1,70	4,270	4,270
Catalyst and Chemical Costs					
Combined Cost, \$/d	351	300	678.5	465	570

* Net consumption.

Table 8-2

(cont'd)

UTILITIES AND CATALYST AND CHEMICAL SUMMARY

Case No.	26	27	28	29	
Claus-Flow Scheme	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru	
-No.of Stgs.	3	3	2	3	
Sulfur Input, t/d	1000	1000	1000	1000	
H ₂ S/CO ₂ Ratio	80/20	80/20	80/20	80/20	
TGCU Process Type	SCOT	BSR/MDEA	BSR/ Selectox	BSR/ Stretford	
Overall S-Rec. %	99.9	99.9	98.6	99.98	
Utilities Required					
Treated BFW, kg/h	153,500	153,500	157,000	154,800	
Electric Power, KW	2,540	2,610	2,590	2,420	
Fuel Gas, m ³ (st)/h	2,130	2,130	2,198	1,295	
Steam, 345 kPa(ga)					
kg/h*	22,500	22,500	-	-	
Cooling Water, m ³ /h	3,370	3,370	362	-	
Utilities Produced					
4140 kPa(ga) steam,					
kg/h	114,200	114,200	114,200	114,200	
345 kPa(ga) steam,					
kg/h	-	-	36,320	36,320	
105 kPa(ga) steam,					
kg/h	-	-	2,190	-	
Steam Condensate,					
kg/h	61,800	61,800	4,270	4,270	
Catalyst and Chemical Costs					
Combined Cost, \$/d	1,234	890	750	1,695	

* Net consumption.

Table 8-3

ANNUAL DIRECT OPERATING COST AND PRODUCT CREDIT SUMMARY

Case No.	1	2	3	4	5
Claus - Flow Scheme	Split Flow	Split Flow	Split Flow	Split Flow	Split Flow
- No. of Stgs.	2	2	2	2	3
Sulfur Input, t/d	50	50	150	150	150
H ₂ S/CO ₂ Ratio	15/85	15/85	15/85	15/85	15/85
TGCU Process Type	CBA	Sulfreen	CBA	Sulfreen	SCOT
Overall S-Rec. %	98.0	98.0	98.0	98.0	99.4
Fixed Capital Cost \$	7,112,000	7,112,000	14,690,000	14,690,000	26,290,000
Direct Operating cost \$ /a					
1. Utilities & Chemicals					
a. Steam	-	-	-	-	65,900
b. Electric Power	76,400	76,400	229,200	229,200	180,500
c. Fuel gas	435,700	435,700	1,307,200	1,307,200	1,747,300
d. Cooling Water	-	-	-	-	118,500
e. Catalyst and Chemicals	20,600	25,900	63,000	80,500	119,000
f. Boiler Fd. Water	187,600	187,600	562,700	562,700	631,800
2. Mainten. & Repair	213,400	213,400	440,700	464,700	789,700
3. Operating Personnel	219,000	219,000	219,000	219,000	219,000
Total Yrly. Cost \$	1,152,700	1,158,000	2,821,800	2,863,300	3,871,700
Credit \$/a					
a. 4140 kPa(ga) steam	496,500	496,500	1,489,500	1,489,000	1,489,000
b. 345 kPa(ga) steam	145,900	145,900	437,700	437,700	-
c. 105 kPa(ga) steam	160,000	160,000	480,000	480,000	248,900
d. Steam Condensate	-	-	-	-	64,400
e. Sulphur	1,208,000	1,208,000	3,624,000	3,624,000	3,664,000
Total Credits \$/a	2,010,400	2,010,400	6,030,700	6,030,700	5,466,300

Table 8-3

(cont'd)

ANNUAL DIRECT OPERATING COST AND PRODUCT CREDIT SUMMARY

Case No.	6	7	8	9	10
Claus - Flow Scheme	Split Flow	Split Flow	Split Flow	Strgt.Thru	Strgt.Thru
- No.of Stgs.	3	2	3	2	2
Sulfur Input, t/d	150	150	150	15	15
H ₂ S/CO ₂ Ratio	15/85	15/85	15/85	50/50	50/50
TGCU Process Type	BSR/MDEA	BSR/ Selectox	BSR/ Stretford	CBA	Sulfreen
Overall S-Rec. %	99.9	97.8	99.98	98.5	98.5
Fixed Capital Cost \$	26,290,000	18,049,000	24,350,000	3,534,000	3,534,000
Direct Operating cost \$ /a					
1.Utilities&Chemicals					
a. Steam	65,900	-	-	-	-
b. Electric Power	186,900	234,400	205,800	18,900	18,900
c. Fuel gas	1,747,200	1,943,700	1,155,300	48,700	48,700
d. Cooling Water	118,500	19,700	-	-	-
e. Catalyst and Chemicals	80,100	105,000	237,500	31,000	3,800
f. Boiler Fd. Water	631,800	667,700	641,800	51,400	51,400
2.Mainten. & Repair	789,700	541,200	730,500	106,000	106,000
3.Operating Personnel	219,000	219,000	219,000	219,000	219,000
Total Yrly. Cost \$	3,839,100	3,730,700	3,189,900	475,000	447,800
Credit \$/a					
a. 4140kPa(ga)steam	1,489,000	1,489,000	1,489,000	151,900	151,900
b. 345 kPa(ga)steam	-	1,016,000	1,016,000	60,700	60,700
c. 105 kPa(ga)steam	248,900	336,400	248,900	9,900	9,900
d. Steam Condensate	233,400	-	-	1,300	1,300
e. Sulphur	3,664,000	3,624,000	3,674,000	360,500	360,500
Total Credits \$/a	5,635,300	6,465,400	6,427,900	584,300	584,300

Table 8-3

(cont'd)

ANNUAL DIRECT OPERATING COST AND PRODUCT CREDIT SUMMARY

Case No.	11	12	13	14	15
Claus - Flow Scheme - No. of Stgs.	Strgt. Thru 2	Strgt. Thru 2	Strgt. Thru 3	Strgt. Thru 3	Strgt. Thru 2
Sulfur Input, t/d	350	350	350	350	350
H ₂ S/CO ₂ Ratio	50/50	50/50	50/50	50/50	50/50
TGCU Process Type	CBA	Sulfreen	SCOT	BSR/MDEA	BSR/ Selectox
Overall S-Rec. %	98.5	98.5	99.9	99.9	98.5
Fixed Capital Cost \$	17,410,000	18,590,000	29,810,000	29,810,000	20,860,000
Direct Operating cost \$ /a					
1. Utilities & Chemicals					
a. Steam	-	-	-	-	-
b. Electric Power	435,500	435,500	405,800	414,200	438,500
c. Fuel gas	1,136,800	1,136,800	1,528,800	1,528,800	1,703,500
d. Cooling Water	-	-	137,500	137,500	3,800
e. Catalyst and Chemicals	65,100	79,800	131,300	94,200	105,000
f. Boiler Fd. Water	1,200,000	1,200,000	1,268,500	1,268,500	1,305,000
2. Mainten. & Repair	522,300	557,700	894,300	894,300	625,800
3. Operating Personnel	219,000	219,000	219,000	219,000	219,000
Total Yrly. Cost \$	3,578,700	3,628,800	4,585,200	4,556,500	4,400,600
Credit \$/a					
a. 4140 kPa(ga) steam	3,542,100	3,542,100	3,542,100	3,542,100	3,542,100
b. 345 kPa(ga) steam	1,414,700	1,414,700	359,500	359,500	1,993,800
c. 105 kPa(ga) steam	230,900	230,900	-	-	87,500
d. Steam Condensate	30,400	30,400	386,200	386,200	30,400
e. Sulphur	8,412,000	8,412,000	8,459,200	8,549,200	8,412,000
Total Credits \$/a	13,630,100	13,630,100	12,837,000	1,837,000	14,065,800

Table 8-3

(cont'd)

ANNUAL DIRECT OPERATING COST AND PRODUCT CREDIT SUMMARY

Case No.	16	17	18	19	20
Claus - Flow Scheme	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru
- No.of Stgs.	3	3	2	2	3
Sulfur Input, t/d	350	100	400	400	400
H ₂ S/CO ₂ Ratio	50/50	80/20	80/20	80/20	80/20
TGCU Process Type	BSR/ Stretford	Not Required	CBA	Sulfreen	SCOT
Overall S-Rec. %	99.98	97.0	98.7	98.7	99.9
Fixed Capital Cost \$	27,720,000	6,200,000	17,230,000	18,330,000	28,630,000
Annual Operating cost, \$					
1.Utilities&Chemicals					
a. Steam	-	-	-	-	1,040,300
b. Electric Power	411,600	94,900	469,100	469,100	466,200
c. Fuel gas	1,004,600	112,600	639,900	639,900	928,200
d. Cooling Water	-	-	-	-	183,800
e. Catalyst and Chemicals	237,500	11,500	65,100	79,800	161,000
f. Boiler Fd. Water	1,279,800	416,900	1,741,000	1,741,000	1,804,500
2.Mainten. & Repair	831,600	186,000	516,900	550,000	858,900
3.Operating Personnel	219,000	131,400	219,000	219,000	219,000
Total Yrly. Cost \$	3,984,100	953,300	3,651,000	3,698,800	5,661,900
Credit \$/a					
a. 4140kPa(ga) steam	3,542,000	1,483,000	6,446,500	6,446,500	6,446,500
b. 345 kPa(ga) steam	1,994,000	334,000	1,101,000	1,101,000	-
c. 105 kPa(ga) steam	-	-	230,900	230,900	-
d. Steam Condensate	30,400	20,300	44,200	44,200	639,000
e. Sulphur	8,573,000	2,377,000	9,604,000	9,604,000	9,770,500
Total Credits \$/a	14,139,400	4,214,300	17,426,600	17,426,600	16,856,000

Table 8-3

(cont'd)

ANNUAL DIRECT OPERATING COST AND PRODUCT CREDIT SUMMARY

Case No.	21	22	23	24	25
Claus - Flow Scheme	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru	Strgt.Thru
- No.of Stgs.	3	2	3	2	2
Sulfur Input, t/d	400	400	400	1000	1000
H ₂ S/CO ₂ Ratio	80/20	80/20	80/20	80/20	80/20
TGCU Process Type	BSR/MDEA	BSR/ Selectox	BSR/ Stretford	CBA	Sulfreen
Overall S-Rec. %	99.9	98.6	99.98	98.7	98.7
Fixed Capital Cost \$	28,630,000	20,330,000	26,890,000	31,810,000	33,240,000
Direct Operating cost, \$					
1.Utilities&Chemicals					
a. Steam	1,040,300	-	-	-	-
b. Electric Power	478,800	473,800	445,200	1,076,500	1,076,500
c. Fuel gas	928,200	959,900	985,200	1,600,000	1,600,000
d. Cooling Water	183,800	19,700	-	-	-
e. Catalyst and Chemicals	102,800	105,000	237,500	162,800	199,500
f. Boiler Fd. Water	1,804,500	1,846,000	1,820,000	4,353,000	4,353,000
2.Mainten. & Repair	858,900	609,900	806,700	954,300	997,200
3.Operating Personnel	219,000	219,000	219,000	262,800	262,800
Total Yrly. Cost \$	5,616,300	4,233,300	4,513,600	8,409,400	8,489,000
Credit \$/a					
a. 4140kPa(ga) steam	6,446,500	6,446,500	6,446,500	16,116,000	16,116,000
b. 345 kPa(ga) steam	-	1,679,500	1,679,500	2,752,000	2,752,000
c. 105 kPa(ga) steam	-	87,500	-	577,500	577,500
d. Steam Condensate	639,000	44,500	44,500	110,500	110,500
e. Sulphur	9,770,000	9,604,000	9,798,000	24,010,000	24,010,000
Total Credits \$/a	16,856,000	17,862,000	17,968,500	43,566,000	43,566,000

Table 8-3

(cont'd)

ANNUAL DIRECT OPERATING COST AND PRODUCT CREDIT SUMMARY

Case No.	26	27	28	29	
Claus - Flow Scheme - No. of Stgs.	Strgt. Thru 3	Strgt. Thru 3	Strgt. Thru 2	Strgt. Thru 3	
Sulfur Input, t/d	1000	1000	1000	1000	
H ₂ S/CO ₂ Ratio	80/20	80/20	80/20	80/20	
TGCU Process Type	SCOT	BSR/MDEA	BSR/ Selectox	BSR/ Stretford	
Overall S-Rec. %	99.9	99.9	98.6	99.98	
Fixed Capital Cost \$	51,740,000	51,740,000	37,040,000	45,230,000	
Direct Operating cost, \$					
1. Utilities & Chemicals					
a. Steam	2,601,000	2,601,000	-	-	
b. Electric Power	1,067,000	1,096,000	1,088,000	1,016,000	
c. Fuel gas	2,326,000	2,326,000	2,400,000	1,414,000	
d. Cooling Water	459,000	459,000	49,000	-	
e. Catalyst and Chemicals	431,900	311,500	262,500	593,200	
f. Boiler Fd. Water	4,513,000	4,513,000	4,616,000	4,551,000	
2. Mainten. & Repair	1,552,200	1,552,200	1,111,200	1,356,900	
3. Operating Personnel	262,800	262,800	262,800	262,800	
Total Yrly. Cost \$	13,212,900	13,121,500	9,789,500	9,193,900	
Credit \$/a					
a. 4140 kPa(ga) steam	16,116,000	16,116,000	16,116,000	16,116,000	
b. 345 kPa(ga) steam	-	-	4,198,000	4,198,000	
c. 105 kPa(ga) steam	-	-	219,000	-	
d. Steam Condensate	1,599,000	1,599,000	110,000	110,000	
e. Sulphur	24,427,000	24,427,000	24,010,000	24,495,000	
Total Credits \$/a	42,142,000	42,142,000	44,653,000	44,919,000	

9.0 REFINERY SULPHUR RECOVERY UNIT INSTALLED AND OPERATING COSTS

In a conventional refinery, there are mainly two acid gas streams sent for sulphur recovery. The first stream contains acid gas from amine plant stripper. This stream is generally rich in H_2S of the order of 80 to 90 mole % and contains a small fraction of COS , CS_2 and RSH . The amount of COS , CS_2 and RSH present in acid gas from a typical refinery does not effect the installed cost, operating cost, and performance of the sulphur plant.

The second stream to sulphur plant is sour gas from sour water stripper. This stream generally contains about 35 mole % NH_3 , about 35 mole % H_2S and 30 mole % water vapors. Complete combustion of NH_3 is required in the sulphur plant reaction furnace to avoid plugging of downstream equipment. Thus, to process the sour gas from sour water stripper, the sulphur plant will have additional features in the reaction furnace and large equipment sizes.

A typical refinery acid gas from amine plant in combination with sour gas from sour water stripper contains 70 - 80 mole % H_2S and 5 - 15 mole % NH_3 . Refinery sulphur plant capacity generally range from 10 to 50 tonnes/day. The capital cost of such a sulphur plant is about 5 to 7% over that to treat an acid gas stream with 80/20 H_2S/CO_2 ratio in the natural gas processing plant as given in Figure 7-1. The operating cost is also about 5 to 7% higher to treat a acid gas stream with 80/20 H_2S/CO_2 ratio in the natural gas processing plant. The operating cost for case No. 17, 100 tonnes/day should be used as a model to arrive at the operating cost for a refinery sulphur plant.

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